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Test and Evaluation of a Polymer Membrane Preconcentrator

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16. ABSTRACT The report gives results of an evaluation of the applicability of membrane systems as a preconcentrator and defines operating parameters of a membrane system. Advantages of such a system is a potential reduction in cost for subsequent control systems. The evaluation is part of a joint EPA/California Air Resources Board investigation of the potential of membrane technology on volatile organic compound (VOC) emissions. Tests of various membrane materials and configurations have been conducted. The polymeric membrane has been used for a number of years as a concentrating step for various liquid and gaseous streams, including the removal of large molecule organics from waste water streams, hydrogen separation, and CO2 recovery. A polymer membrane is an ultrathin layer of a selective polymer, supported on a porous sublayer. The membrane (active layer) selectively filters the pollutant molecules. A potentially innovative application of membrane technology may be to concentrate VOCs from exhaust gases such as solvent oven-drying exhaust. A preconcentrator membrane could be used to reduce the size and, in turn, the capital and operating costs of a conventional VOC control device such as a carbon adsorber or incinerator. Study results do not, however, verify that a membrane preconcentrator is a viable option to reduce overall pollution control costs.

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TEST AND EVALUATION OF A POLYMER MEMBRANE PRECONCENTRATOR

FINAL REPORT

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Abstract

The control of emissions of Volatile Organic Compound emissions to the atmosphere from industrial operations represents one of the most challenging problems in the area of pollution control. Although technologies are available to capture or destroy these emissions, the cost in many cases has proven prohibitive. The utilization of polymeric membranes may represent an opportunity to reduce the cost of controlling these emissions. Polymeric membranes have been used for a number of years as a concentrating step for various liquid and gaseous streams, including the removal of large molecule organic from waste water streams, hydrogen separation and CO_2 recovery. A polymer membrane is an ultra-thin layer of a polymer material which has the capability of selectively filtering the polluting molecule.

This report gives results of research on the applicability of membranes to enhance the ability of existing control technologies to capture or destroy these emissions. The report evaluates membranes and defines operating parameters of a membrane system. The research project is a joint effort between the EPA and the California Air Resources Board. Test of various membrane materials and configurations have been conducted.

The potentially innovative application of membrane technology may be to concentrate the polluting VOC in the process exhaust reducing its volume thus permitting more economical control of the pollutant. Due to the projected high cost of operating the membrane, however, this study could not draw the conclusion that it is a viable option to enhance pollution control.

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ABBREVIATIONS AND ACRONYMS

ACFM	Actual cubic feet per minute
AMCEC	American Ceca Corporation
ARB	California Air Resources Board
Atm	Atmospheres
CA	Chemical Abstracts
CFC-113	1,1,2-trichloro-1,2,2-trifluoroethane
Compendex	Engineering Index Computer Database
CO_2	Carbon dioxide
DGM	Dry gas meter
DQIs	Data Quality Indicators
DQOs	Data Quality Objectives
EPA	U.S. Environmental Protection Agency
F	Stage cut
HCl	Hydrogen chloride
HF	Hydrogen fluoride
HP	Horsepower
INEL	Idaho National Engineering Laboratory
kWh	Kilowatt-hours
LEL	Lower Explosive Limit
MeCl	Methylene Chloride
MEK	Methyl Ethyl Ketone
MeOH	Methyl Alcohol (Methanol)
mmHg	Millimeters of mercury
MT	Metric Tons
MTR	Membrane Technology and Research
NBS	National Bureau of Standards (now NIST)
Nm^3	Normal cubic meters
NTIS	National Technical Information Service
PI	Pressure indicator
ppmv	Parts per million by volume
psia	Pounds per square inch absolute
P_T	Total pressure

QA	Quality assurance
QAPP	Quality Assurance Project Plan
QC	Quality Control
QCER	Quality Control Evaluation Report
r	Linear correlation coefficient
RF	Response factor
RH	Relative humidity
SCFM	Standard cubic feet per minute
SLPM	Standard liters per minute
SS	Stainless steel
STP	Standard temperature and pressure (0°C and 760 mmHg)
THC	Total hydrocarbons
TI	Temperature indicator
TSA	Technical Systems Audit
UHP	Ultra High Purity
VOC	Volatile Organic Compound

SECTION 1

INTRODUCTION

The California Air Resources Board (ARB) and the U.S. Environmental Protection Agency (EPA) both seek to identify new and innovative methods to control toxic air pollutants. One potentially viable concept is the use of polymeric membrane materials which allows the selective permeation of organic vapors. However, further development and testing will be required before this technology can be considered as a proven near-term solution.

OVERVIEW OF THE TECHNOLOGY

Membrane systems have been used for several years as a concentrating step for various operations such as water treatment, hydrogen separation, and CO_2 recovery. A polymeric membrane system for organic vapor recovery typically consists of an ultra-thin layer of a selective polymer which is supported on a porous sublayer (see Figure 1-1). The open support material is used as a spacer to separate the polymer layers in a spiral-wound membrane module.

An innovative use of a membrane may be for concentrating hydrocarbon vapors from exhaust gases such as solvent oven drying exhaust. A "preconcentrator" membrane could be used to reduce the size and, in turn, the capital and operating requirements of a conventional VOC control device such as a carbon adsorber or incinerator. The overall result would be a cost savings, a performance improvement (i.e., greater emissions reductions), and, for incinerators, reduced energy requirements.

The purpose of this work is to evaluate the applicability of membrane systems as a preconcentrator and to define operating parameters of a membrane system. The advantages of such a system are a potential reduction in cost for the overall system both from a capital and operating cost standpoint and a

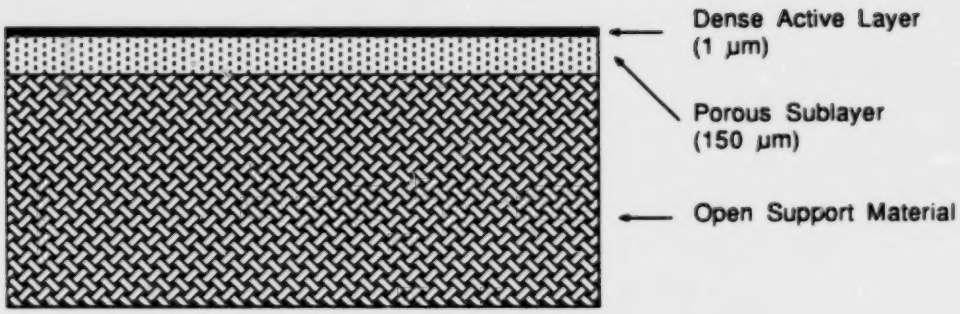

Figure 1-1. Schematic Cross-Section of a Composite Polymer Membrane

potential increase in the number of applications that could use those conventional controls, both technically and economically.

In order to achieve the objectives, several tasks were performed. First, a bench-scale membrane module was tested with six common solvents to define the capability of membrane technology to solve toxic air emission problems and to define operating parameters. Next, the bench-scale data was used to develop preliminary conceptual system designs. With these designs, cost estimates were prepared for both capital and operating costs of the membrane assisted systems, and these costs were compared to the costs for systems which did not utilize the membrane preconcentrator step.

As a prelude to the experimental work, a review of available literature on hydrocarbon vapor recovery with membranes was performed. Any relevant articles found in the search are discussed below.

LITERATURE REVIEW

In order to obtain any additional information on membranes, specifically gas-phase hydrocarbon recovery, a computerized literature search was performed. Unfortunately, very little new data were found. Out of three major databases (National Technical Information Service (NTIS), Chemical Abstracts (CA), and Engineering Index (COMPENDEX)), only seven entries were located, three of which were not applicable. The conclusion is that membrane applications in the VOC recovery area are rare.

Theoretical studies have been presented for polymeric membrane systems. The fundamental material and energy balance equations governing the design and performance of single-stage gas permeation were presented by Weller and Steiner (1). A further analysis for the cross-flow pattern (which applies to the spiral-wound module used in this study) was performed by Pan and Habgood (2). The theoretical model in Section 4 used for comparison with the experimental data is based entirely on the equations found in Pan and Habgood.

Much of the data on gas-phase hydrocarbon recovery using polymeric membranes have been presented by Membrane Technology and Research, Inc. (MTR). The effort at MTR has been led by R.W. Baker, and has resulted in several papers (3,4), reports (5,6), and at least one patent (7). MTR is actively marketing a membrane system for solvent vapor recovery for smaller industrial applications such as web drier emissions.

A recent paper (8) dealt with synthetic membranes for separation of organic vapors from waste air streams. The authors discuss their tests of hollow fiber membranes using polydimethylsiloxane as the selective barrier. They propose a process for recovery of toluene from spray painting operations.

SECTION 2

SUMMARY AND CONCLUSIONS

SUMMARY

A bench-scale polymeric membrane system was designed and constructed for this program. The membrane was spiral-wound and was supplied by a current membrane manufacturer. The membrane test module performed well in removing a large percentage of solvent from dilute (20 to 2000 ppmv) gas streams. The membrane was able to remove about 60 percent of the incoming solvent, and generated a "permeate" stream about three (3) times as concentrated as the original feed. The module was equally effective on all six of the solvents tested. No noticeable degradation in performance of the module was apparent after the test sequence, although an extended performance evaluation was not conducted.

Based on the test data and available cost data for two simple configurations, the membrane preconcentrator does not appear to be an economic alternative to carbon adsorption for low concentration (i.e., 100 to 1000 ppmv) solvent-laden air streams. The capital and annualized costs of the membrane-augmented system were consistently higher than the carbon adsorber alone. Cost reductions for the membrane-augmented carbon adsorber (due to the reduced volume flow) were not sufficient to cover the added expense of the membrane and associated equipment.

Additionally, the study examined the use of a pressurized feed versus a vacuum permeate stream. For equivalent inlet gas flows, the vacuum-pumped arrangement was more expensive than the compressed feed arrangement. This was surprising since it seems wasteful to compress the full feed flow rather than the smaller permeate flow. Nevertheless, the compressed feed arrangement requires less membrane area, and avoids potential problems with humidification of the permeate when using a liquid ring vacuum pump.

CONCLUSIONS

The test program was able to provide reproducible data regarding the performance of the bench-scale membrane module. The sampling and analytical methods worked well, and the data could be correlated to an existing model. The program was able to characterize the operation of membrane module.

The conceptual design phase of this project provided an opportunity to study the material balance equations developed by Weller and Steiner (2) for cross-flow (spiral-wound) membranes. The material balance model was able to accurately approximate the experimental performance data for stage cut and enrichment ratio. Discussions with carbon adsorber vendors brought out additional design considerations, especially regarding the upper limit for enrichment (25 percent of the lower explosive limit for flammable solvents), and the potential problems of saturating the permeate stream with water vapor when a water-sealed liquid ring vacuum pump is used.

The cost algorithm developed for a membrane-augmented system showed it to be more costly than direct carbon adsorption in all of the cases studied. The cost estimations predicted that capital costs of carbon adsorbers do not change much at flow rates below 28.3 Nm^3/min (1000 scfm). Furthermore, the benefit of increased inlet concentration to the adsorber (i.e., reduced volume flow) is not great, since the amount of carbon in the bed (and the amount of steam required) is dependent on the amount of solvent to be handled. Thus, until the working capacity of the membrane system is significantly improved and operating cost reduced, this technology cannot be considered economical for pollution control.

SECTION 3

EXPERIMENTAL TESTING

TEST OBJECTIVES

The purpose of the bench-scale testing was to obtain experimental data on the performance of a small spiral-wound membrane module used to concentrate solvent vapors. In past reports by others, much of the experimental data were obtained using very small permeation cells containing a flat membrane disc of only a few square centimeters in area. Extrapolation of experimental data from such a small membrane is highly uncertain. Instead, this study has used a small spiral-wound membrane module to obtain data which may be scaled up with more confidence. The approximate membrane area of the test module is 0.4 m^2 (4.3 ft^2). Other information about the membrane module is presented in Table 3-1.

The membrane performance is indicated by two properties: 1) the removal efficiency, or in other words, the percentage of solvent entering which is transferred to the permeate stream (related to the solvent flux across the membrane); and 2) the separation factor, which is the degree of concentration or enrichment which the membrane can achieve (related to the selectivity of the membrane). Both of these properties are dependent on the operating conditions. For example, the pressure ratio (permeate-side pressure/inlet pressure) can exert a strong influence on removal efficiency and enrichment.

The experimental tests were conducted on six solvents (listed in Table 3-2). Each of the tested solvents finds wide use in commercial and industrial applications and is meant to represent certain classes of organics. Also, the solvent vapor feed concentration was varied for each of the solvents, generally within the range of 20 to 2,000 parts per million volume (ppmv). This low concentration range was chosen since data in this range has not been available in the literature. Also, it is the range where the membrane system

TABLE 3-1. MEMBRANE MODULE DATA

Manufacturer:	Nitto Electric Industrial Co., Ltd. Shiga Plant, Membrane Division Kusatsu, Shiga, Japan
Module Type:	S2B Organic Vapor Recovery Module
Configuration:	Spiral Wound
Membrane Material:	Composite Polyimide
Membrane Area:	0.4 m^2 (4.3 ft^2)
Membrane Thickness:	3μm (.0012 in)
Module Dimensions:	(External Housing) 7.9 cm diameter x 62.0 cm length (Internal Element) 6.1 cm diameter x 53.3 cm length
Normal Operating Conditions:	Inlet Flow: 50 SLPM (2.0 scfm) Inlet Pressure: 800 mmHg (15.5 psia) Permeate Flow: 3 SLPM (0.106 scfm) Permeate Pressure: 80 mmHg (1.55 psia) Temperature: 25°C (77°F)

TABLE 3-2. SOLVENTS TESTED IN THE STUDY

Solvent	Class	Industrial Application
Hexane	Aliphatic	Surface coating Vegetable extraction
Toluene	Aromatic	Surface coating Printing
MEK[a]	Ketone	Surface coating Printing
Methanol	Alcohol	Printing Degreasing
Freon 113[b]	Chlorofluorocarbon	Electronic degreasing Dry cleaning
Methylene Chloride	Chlorinated	Metal degreasing Foam blowing

[a]Methyl ethyl ketone
[b]1,1,2-Trichloro-1,2,2-trifluoroethane

may be applied as a preconcentrator in conjunction with other VOC control technologies which work best on high solvent concentration streams. Test conditions that were measured and kept constant included: 1) inlet temperature, pressure, and flowrate; 2) outlet pressure and flowrate; and 3) permeate pressure. The module was tested with solvent vapor in nitrogen gas. This reduced the potential hazards associated with several of the flammable solvents and also provided data comparable to previous studies in nitrogen.

The flammability of certain solvents could pose an additional hazard to the testing or practical application of membrane systems for solvent recovery. Flammable solvents and oxygen can form explosive mixtures. A membrane preconcentrator handling (flammable) solvent vapors in air could result in shifting the mixture from a dilute feed condition below the lower explosive limit (LEL) to a concentrated permeate mixture within the explosive range.

TEST PROCEDURES

Figure 3-1 illustrates the experimental apparatus. The membrane module was installed in a closed-loop arrangement for testing. The closed-loop arrangement was chosen to minimize the amount of nitrogen which would otherwise be wasted if the system vented to the atmosphere. The stripped off-gas or residue was recycled back to the inlet and a small volume of make-up solvent vapor was fed to the loop. Because of the small (but finite) amount of nitrogen which passed through the membrane with the solvent, make-up nitrogen from a cylinder was required.

The primary measurements were the total hydrocarbon (THC) concentrations taken at the membrane feed inlet and stripped off-gas outlet. These concentrations were measured with the Byron 401 THC analyzer, with samples taken semi-continuously at one-minute intervals. Since the project utilized only one analyzer, it was necessary to sample the inlet and outlet locations alternately.

In order to close a material balance around the system, it was necessary to measure the amount of solvent transferred to the permeate stream. During

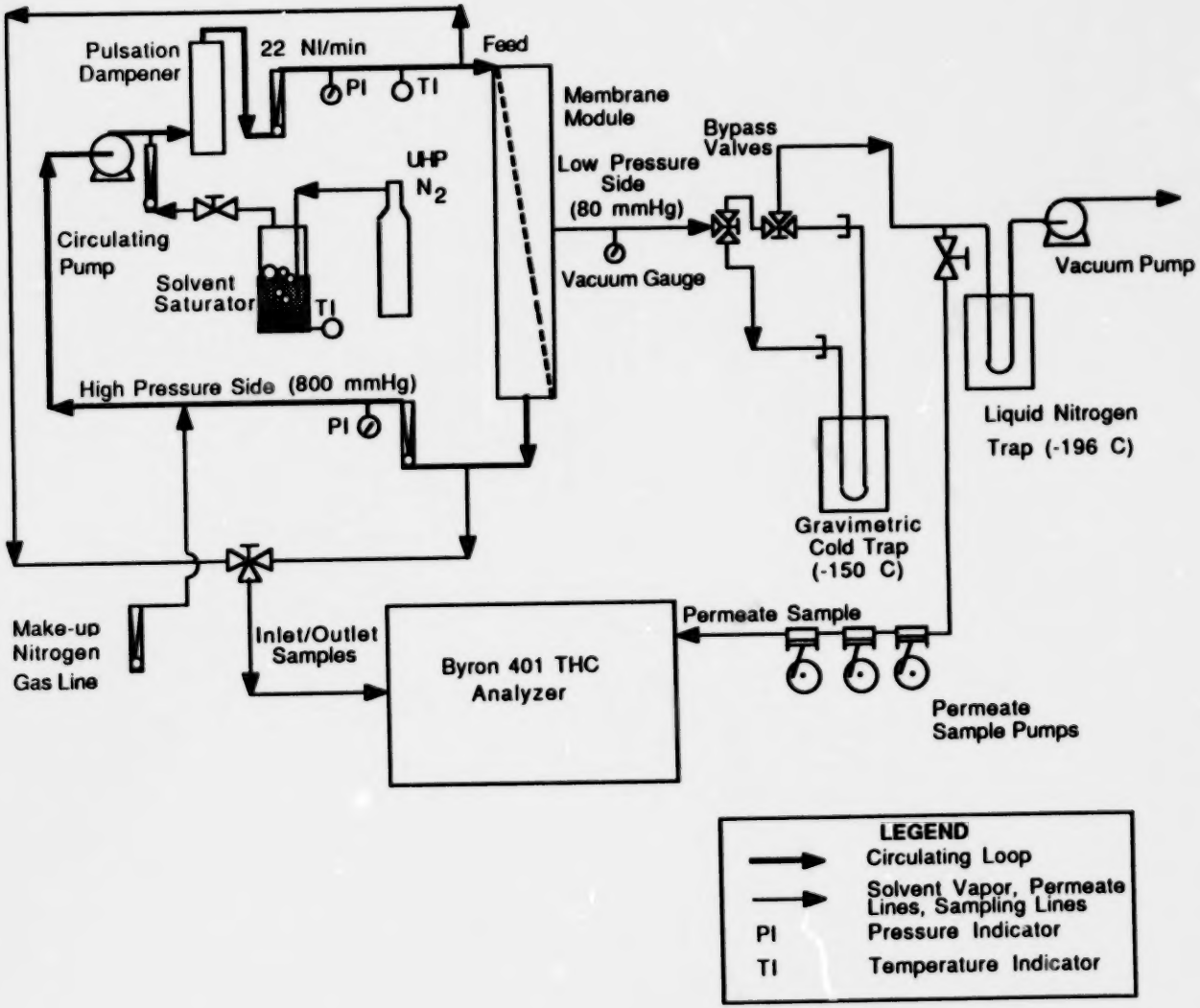

Figure 3-1. Schematic Diagram of Test Apparatus

the course of the project, tests showed that the cold trap sampling technique was not sufficiently accurate to determine the permeate solvent flux. As explained later, a method was developed to extract a sample from the permeate stream for direct THC analysis. This required a special arrangement of additional sample pumps (operated in series) to pull samples from the low pressure permeate side (see Figure 3-1).

Starting a test series on a new solvent first involved filling the previously-cleaned saturator with Reagent Grade solvent and closing the top. A Teflon® gasket sealed the flanged connection. The saturator used a dip tube to bubble nitrogen through the solvent. The saturator also contained 1/2" glass Raschig rings (packing) to ensure adequate gas-liquid contacting. The temperature of the liquid solvent was measured with a thermocouple. The saturator was wrapped with a heating tape (and insulation) to allow the solvent to be heated. The saturator was pressurized to a known pressure. By varying the solvent temperature and saturator pressure, it was possible to adjust the concentration of the saturated solvent vapor.

Next, the membrane system was started with nitrogen only to establish a steady-state flow and total hydrocarbon background before adding saturated solvent vapor to ensure the removal of traces of solvent from previous tests. A low baseline level of solvent was determined by sampling the inlet THC concentration with the Byron 401. The circulating pump and vacuum pump were started before beginning the flow of saturated solvent vapor into the circulating loop. Normally, the cold trap was filled with liquid nitrogen, and the vacuum was adjusted as the circulating pump was started. After establishing a low baseline for inlet THC, the flow of saturated solvent vapor could be started. The solvent concentration in the circulating loop was governed mostly by the flowrate of the solvent vapor.

Once the inlet THC concentration had stabilized, data collection was started. Sampling the inlet and outlet THC (and later, permeate THC) streams involved switching back and forth, since only one THC analyzer was used. Confidence in the data was highest when the respective THC values remained essentially constant between sampling periods. Whenever possible, the THC

sampling sequence was from low to high concentration. This helped to minimize delays and potential erroneous THC responses caused by solvent adsorption on the Teflon sample lines. Therefore, sample lines were changed when switching from the permeate (highest concentration) to the outlet (lowest concentration). If contaminated, the sample line could be cleared in a short time by allowing UHP N_2 to flow through it.

Overall, including the daily multipoint calibration of the THC analyzer, it was possible to complete tests at two levels of inlet solvent concentration each day. This is based on the time required to change solvents, purge the membrane, and collect about two hours of data at each condition.

QA/QC Procedures

Several types of procedures were developed to provide quality assurance (QA) and quality control (QC). These QA/QC procedures were detailed in the Test Plan/Quality Assurance Project Plan (QAPP). A brief list of these procedures is summarized in Table 3-3. The QAPP was written to ensure that the experimental measurements would provide results of sufficient quality to evaluate the performance of the control technology. The QAPP discussed Data Quality Objectives (DQOs), Data Quality Indicators (DQIs), sampling and analytical procedures, data reduction methods, data validation methods, and reporting procedures. The statistical analysis of the test data is presented in the Quality Control Evaluation Report (QCER), in Appendix E of this report. The detailed results of the calibrations and other QC checks are presented in Appendix B.

TABLE 3-3. QUALITY CONTROL CHECKS

Parameter	Method of Measurement	Type of QC Check	Frequency	Standards	Acceptance Criteria
VOC	Byron 401 THC Analyzer	Multipoint (Propane)	Weekly	Propane in N_2	$r \geq 0.995$
VOC	Byron 401 THC Analyzer	Multipoint (Solvent)	Daily	Certified Master Gas of Solvent in N_2	$r \geq 0.995$
VOC	Gravimetric Trap (Electronic Top Loading Balance)	Multipoint	Weekly	Class S weights	±2%
VOC	(Gravimetric Trap/ Direct Perm Sampling)	Recovery Test	Once	Material Balance	>95%
Flowrate	Rotameter	Multipoint	Once	Hastings Raydist Flow Calib. or Calibrated DGM	±1%
Temperature	Thermocouple Type J (Saturator) Type K (Inlet Gas)	Multipoint	Once	NBS Calib. Thermometers	±1%
Pressure	Magnehelic	Multipoint	Once	Inclined Manometer	±1%
Pressure/Vacuum	Bourdon Tube Gauge	Multipoint	Once	Reference Test Gauges	±1%
Barometric Pressure	Mercury Manometer	-	Daily	National Weather Service Office	-

SECTION 4

TEST RESULTS

EXPERIMENTAL DATA

Experimental data were recorded in a laboratory notebook and on a strip chart recorder connected to the Byron 401 THC analyzer. At the conclusion of the experimental testing, the laboratory notebook data were entered into a PARADOX relational database for further data manipulation. Appendix A contains a tabulated listing of process data taken (approximately) every five minutes. Also included in Appendix A are sample calculations which illustrate how the process data were converted to the intermediate results of solvent mass flow. Appendix E contains the statistical analysis of the results, including estimates of experimental error. Table 4-1 presents a summary of the average inlet, outlet, and permeate concentrations for each test. Also listed are average inlet and outlet solvent mass flowrates and the average pressure ratio of each test. For the test runs where the gravimetric trap was used, the trap results were not used, and the remaining calculations for those runs were based on an assumption of 100% closure of the material balance (a more detailed discussion is provided in a subsequent section).

REMOVAL EFFICIENCY

Removal efficiency refers to the percentage of incoming solvent vapor which is removed by the membrane. For example, consider a source of solvent-laden exhaust air containing 100 kg/hr of solvent. If a membrane having a removal efficiency of 75% were to be applied to this air stream, then the membrane would produce a stripped off-gas containing only 25 kg/hr of solvent. The balance (75 kg/hr) would be contained in the more concentrated permeate stream.

The removal efficiency of the membrane module was calculated by the following equation:

$$\text{Removal Efficiency} = \frac{(\text{Inlet Solvent Flow} - \text{Outlet Solvent Flow}) \times 100\%}{\text{Inlet Solvent Flow}} \qquad (4\text{-}1)$$

The numerator (Inlet - Outlet) is equal to the solvent flux through the membrane. Results of removal efficiency for each test are shown in Table 4-1.

ENRICHMENT RATIO

Enrichment ratio refers to the degree of enrichment that the membrane can accomplish at given conditions. For example, consider a source of solvent-laden exhaust air with an initial concentration of 1000 ppmv. If a membrane having an enrichment ratio of five were applied to this air stream, then the membrane would produce a permeate stream enriched to 5000 ppmv.

The enrichment ratio is simply the ratio of the permeate concentration to the inlet (feed) concentration. Table 4-1 presents the average enrichment ratio for each test.

SEPARATION FACTOR

Separation factor refers to the relative permeabilities of the solvent and the gas (e.g., nitrogen or air) through the membrane. If the separation factor were equal to one, then the permeabilities would be equal and no separation could be obtained. From the standpoint of trying to optimize a membrane for high removal and enrichment of solvent, one would prefer a high separation factor. However, as will be discussed later, membrane area is not determined by the relative permeability, but by the actual permeability of the solvent through the membrane. Therefore, separation factor is usually a compromise.

The separation factor (alpha) was calculated by the following equation:

TABLE 4-1. SUMMARY OF AVERAGE TEST RESULTS

Solvent	Test Date	Inlet Solvent Concentration (ppmv)	Inlet Solvent Mass Flow (g/hr)	Outlet Solvent Concentration (ppmv)	Outlet Solvent Mass Flow (g/hr)	Removal Efficiency (%)	Permeate Solvent Concentration (ppmv)	Enrichment Ratio	Pressure Ratio
Hexane	12-14-88	327	1.65	121	0.50	69.8	1260	3.86	0.046
	12-14-88	1518	7.74	556	2.31	70.1	5770	3.80	0.055
	01-20-89	64.6	0.33	31.5	0.13	59.9	189	2.93	0.075
	01-20-89	716	3.66	360	1.54	58.1	2210	3.09	0.075
Toluene	01-19-89	33.0	0.179	15.6	0.069	61.2	100	3.04	0.072
	01-19-89	128.0	0.69	56.5	0.25	63.6	409	3.19	0.072
MEK	12-16-88	60.6	0.26	17.7	0.063	76.0	260	4.30	0.049
	12-19-88	346	1.48	166	0.59	60.2	1220	3.52	0.075
	01-03-89	1725	7.43	770	2.74	62.7	6220	3.61	0.087
MeOH	01-13-89	84.5	0.16	54.2	0.086	47.0	299	3.53	0.072
	01-13-89	376	0.71	192	0.30	57.7	1260	3.35	0.072
	01-16-89	1190	2.17	582	0.87	59.8	4000	3.37	0.072
CFC-113	01-09-89	16.8	0.186	11.4	0.103	44.3	41	2.47	0.088
	01-09-89	148	1.64	83.5	0.76	53.6	442	2.99	0.086
	01-10-89	1371	15.2	684	6.27	58.8	4630	3.38	0.084
MeCl	01-05-89	91.5	0.45	44.6	0.18	60.0	306	3.35	0.074
	01-06-89	88.7	0.44	38.3	0.16	64.5	259	2.92	0.072
	01-06-89	1257	6.24	578	2.35	62.3	4336	3.45	0.072

$$\text{Separation Factor} = \frac{\text{Permeability of Solvent}}{\text{Permeability of Nitrogen}} \qquad (4\text{-}2)$$

where, Permeability = Flux/partial pressure difference;
Flux = Inlet flow per unit time; and
Partial Pressure Difference = Inlet side partial pressure - permeate side partial pressure.

Table 4-2 presents the average values used to calculate the separation factor for each run. Solvent permeabilities are functions of the solvent feed concentration and solvent type (8). Therefore, it is not unexpected that the separation factors differ between runs.

MATERIAL BALANCE

In order to check the validity of the inlet and outlet solvent flow measurements, a material balance was performed around the membrane module. A material balance can be shown mathematically as:

$$\text{IN} - \text{OUT} + \text{ACCUMULATION} = 0 \quad \text{(Conservation of Mass)} \qquad (4\text{-}3)$$

or, for steady state conditions with no accumulation:

$$\text{IN} = \text{OUT}$$

or

$$\text{OUT/IN} = 100\%$$

where,

IN = Inlet Solvent Mass Flow (g/hr)
OUT = Permeate + Outlet Solvent Mass Flow (g/hr)

The material balance used an independent measurement of permeate concentration (or permeate flux) for comparison with the solvent flux used in the calculation for removal efficiency. The following discussion concerns the results of two different sampling techniques which were used to obtain this data.

TABLE 4-2. SEPARATION FACTOR RESULTS

Solvent	Test Date	Inlet Pressure (mmHg)	Permeate Pressure (mmHg)	SOLVENT Partial Pressure Difference	SOLVENT Flux Rate (L/day*m^2)	SOLVENT Permeability (gmol/hr*mmHg)	NITROGEN Partial Pressure Difference	NITROGEN Permeability (gmol/hr*mmHg)	Separation Factor (α)
Hexane	12-14-88	777.8	35.4	0.130	0.105	0.103	742.2	0.014	7.246
	12-14-88	778.5	43.0	0.559	0.494	0.113	734.9	0.015	7.641
	01-20-89	778.9	58.3	0.026	0.016	0.087	720.6	0.014	5.991
	01-20-89	779.2	58.3	0.290	0.172	0.085	720.6	0.014	6.205
Toluene	01-19-89	776.8	56.3	0.013	0.007	0.090	720.5	0.014	6.075
	01-19-89	777.0	56.3	0.049	0.027	0.098	720.7	0.014	6.880
MEK	12-16-88	792.2	38.7	0.021	0.015	0.132	753.5	0.01	9.389
	12-19-88	778.9	58.3	0.129	0.066	0.096	720.5	0.01	6.847
	01-03-89	789.1	68.3	0.560	0.350	0.116	720.3	0.01	8.073
MeOH	01-13-89	777.1	56.3	0.037	0.007	0.064	720.8	0.01	4.460
	01-13-89	776.8	56.3	0.150	0.031	0.085	720.4	0.01	6.109
	01-16-89	777.03	56.3	0.462	0.103	0.088	720.3	0.01	6.185
CFC-113	01-09-89	790.0	69.3	0.008	0.003	0.054	720.7	0.01	3.618
	01-09-89	790.2	67.9	0.061	0.034	0.077	722.2	0.01	5.211
	01-10-89	786.1	66.2	0.501	0.340	0.095	719.3	0.01	6.684
MeCl	01-05-89	778.3	57.3	0.035	0.012	0.091	721.0	0.01	6.236
	01-06-89	777.1	56.3	0.035	0.011	0.096	720.8	0.01	6.642
	01-06-89	777.1	56.3	0.469	0.176	0.098	720.4	0.01	6.685

Gravimetric Trap

The original approach to determine the permeate solvent flux was to use a pre-weighed cold trap to condense the permeate solvent vapor over a known time period. The difference between the post-test weight and the tare weight would yield the amount of solvent collected. This technique was tested prior to conducting the actual experiments (Appendix B), and was used during the test runs before the apparatus was modified for direct permeate sampling. Unfortunately, the QA/QC tests showed that the recovery of the cold trap was lower than expected. As a result, the isopentane/liquid nitrogen bath was changed to liquid nitrogen in an attempt to improve sample recoveries by maintaining the trap at an even lower temperature. However, as seen in Table 4-3, the closures with the gravimetric trap were consistently low. We speculate that the cause is incomplete solvent trapping caused by poor heat transfer in the trap and a short residence time.

Direct Permeate THC Sampling

An alternative approach to obtaining an independent value for the permeate flux involved pulling a sample from the permeate stream and analyzing it with the THC analyzer. This required using a two-stage Thomas pump and a separate diaphragm pump to obtain the necessary vacuum. This technique was also tested prior to collecting experimental data; the results are presented in Appendix B. As Table 4-4 indicates, material balance closures obtained with this sampling method are much closer to 100%.

COMPARISON WITH THEORETICAL MODEL

When prior experience with full-scale systems is lacking, many engineering studies use pilot-scale or bench-scale results to assist in the design of full-scale units. The accepted basis for "scale up" estimates is the use of dimensionless groups which are derived from the application of the laws of conservation of mass, momentum, and energy.

TABLE 4-3. CLOSURES WITH GRAVIMETRIC TRAP

Test Date	Time	Solvent	Inlet Solvent Mass Flow (g/hr)	Outlet Solvent Mass Flow (g/hr)	Solvent Captured in Trap (g/hr)	Percent Closure
12-14-88	4:10-5:10	Hexane	1.65	0.50	0.46	58.18%
12-14-88	7:30-8:30	Hexane	7.74	2.31	3.04	69.12%
12-16-88	4:00-5:30	MEK	0.262	0.063	0.587	248.1%
12-19-88	8:20-9:46	MEK	1.482	0.591	0.419	68.15%
01-03-89	8:34-10:24	MEK	7.434	2.741	2.01	63.91%
01-09-89	7:04-8:11	CFC-113	1.64	0.76	0.412	71.46%
01-10-89	10:48-11:32	CFC-113	15.204	6.269	6.395	83.29%
01-10-89	11:42-12:22	CFC-113	15.204	6.269	4.905	73.49%
01-16-89	2:15-4:10	MeOH	2.174	0.874	0.60	67.8%

TABLE 4-4. CLOSURES WITH DIRECT PERMEATE SAMPLING

Test Date	Time	Solvent	Inlet Solvent Mass Flow (g/hr)	Outlet Solvent Mass Flow (g/hr)	Permeate Solvent Mass Flow (g/hr)	Percent Closure
01-20-89	5:00-6:30	Hexane	0.33	0.13	0.17	91.78%
01-20-89	6:50-7:50	Hexane	3.66	1.54	1.89	93.50%
01-06-89	2:45-4:40	MeCl	0.437	0.155	0.238	95.75%
01-06-89	5:30-6:50	MeCl	6.235	2.347	3.908	98.69%
01-13-89	2:10-4:30	MeOH	0.162	0.086	0.098	114.4%
01-13-89	5:25-6:50	MeOH	0.709	0.300	0.403	99.12%
01-16-89	2:30-5:00	MeOH	2.174	0.874	1.353	99.81%

The analysis of a single-stage gaseous permeation process was performed by Pan and Habgood (2). They expanded on previous work by Weller and Steiner (1). Among the simplifying assumptions are: permeabilities of both components are constant; negligible pressure drop across feed and permeate flow paths; and, negligible mass transfer resistances other than the permeation process itself. As mentioned earlier, the assumption of constant permeability is not strongly supported by the experimental data, nor by other researchers (8). However, examination of the results in Table 4-2 shows that the error introduced by this approximation is not excessive in most cases. The test procedures were not sophisticated enough to provide data to check the other assumptions.

The spiral-wound membrane module used in this study follows the cross-flow pattern. It has been shown that as the feed concentration approaches zero, the equations used to describe membrane performance simplify considerably (2). The equations are shown below:

$$F = 1 - (x/x_f)^{[\alpha^*/(1-\gamma)(\alpha^*-1)]-1} \qquad (4\text{-}4)$$

$$y/x_f = [1 - (x/x_f)^{\alpha^*/(1-\gamma)(\alpha^*-1)}]/F \qquad \text{("enrichment ratio")} \qquad (4\text{-}5)$$

$$R^f = F/(1-\gamma) \qquad (4\text{-}6)$$

where, F = fraction of feed permeated in cross-flow pattern ("stage cut");

x_f = mole fraction of solvent in feed gas;

x = mole fraction of solvent in residue (off-gas); and

$\alpha^* = Q_S/Q_{N2}$ = permeability of solvent/permeability of nitrogen ("selectivity").

Q_s = Permeability of solvent $\quad \dfrac{\text{ml(STP)} \cdot \text{cm}}{\text{cm}^2 \cdot \text{sec} \cdot \text{mmHg}}$

$\gamma = p/P$, permeate/feed pressure ratio

Y = mole fraction of solvent in permeate side stream (average permeate concentration in cross-flow pattern)

$R^f = (Q_{N2}/d)PS/L_f$ = dimensionless membrane area with references at feed inlet end

d = membrane thickness, ft

P = feed side pressure, psia

S = membrane area, ft^2

L_f = inlet feed flowrate, lb•mol/hr

These equations are valid for finite $1/x_f$ provided that $1/x_f$ is greater than both α^* and $1/(\gamma)$. Both of these inequalities are valid for these experiments.

Stage cut (F) represents the fraction of feed gas which passes through the membrane into the permeate. While this definition sounds similar to removal efficiency, it is different. Recall that removal efficiency referred only to the percentages of solvent which was removed, while stage cut refers to the fraction of total feed gas removed (i.e., solvent and nitrogen). Therefore, removal efficiency and stage cut, while similar, are nonetheless different quantities.

As shown in Table 4-5, experimental enrichment ratios show good agreement with the calculated values. Also, comparison of calculated stage cuts (F) with experimental stage cuts showed good agreement.

Thus, we conclude that the good agreement of both experimental enrichment ratios and stage cuts with the model indicates a) the material balance results are accurate, and b) there is justification in using the model to confidently extrapolate membrane designs.

TABLE 4-5. COMPARISON BETWEEN EXPERIMENTAL STAGE CUT AND ENRICHMENT RATIO WITH THEORY

Solvent	Test Date	Average Inlet Conc. (ppm)	Stage Cut		Enrichment Ratio	
			Exper.	Theory	Exper.	Theory
Hexane	12-14-88	327	0.181	0.193	3.86	3.64
Hexane	12-14-88	1518	0.184	0.197	3.80	3.59
Hexane	01-20-89	64.6	0.176	0.193	2.93	3.15
Hexane	01-20-89	716	0.167	0.180	3.09	3.27
Toluene	01-19-89	33.0	0.181	0.195	3.04	3.17
Toluene	01-19-89	128	0.175	0.192	3.19	3.34
MEK	12-16-88	60.6	0.176	0.195	4.30	3.91
MEK	12-19-88	346	0.171	0.177	3.52	3.41
MEK	01-03-89	1725	0.175	0.182	3.61	3.48
MeOH	01-13-89	84.5	0.173	0.159	3.53	2.89
MeOH	01-13-89	376	0.171	0.177	3.35	3.28
MeOH	01-16-89	1188	0.180	0.185	3.37	3.25
CFC-113	01-09-89	16.8	0.181	0.182	2.47	2.45
CFC-113	01-09-89	148	0.179	0.183	2.99	2.94
CFC-113	01-10-89	1371	0.174	0.179	3.38	3.29
MeCl	01-05-89	91.5	0.179	0.186	3.35	3.25
MeCl	01-06-89	88.7	0.178	0.203	2.92	3.24
MeCl	01-06-89	1257	0.181	0.188	3.45	3.34

SECTION 5

CONCEPTUAL DESIGNS OF MEMBRANE PRECONCENTRATOR

PROCESS DESCRIPTION

The membrane module is one component in a organic vapor recovery system that includes several other elements. A simplified overall diagram of a membrane system for solvent vapor recovery is shown in Figure 5-1. Organic solvent vapors generated by the source (such as a solvent degreaser machine or a drying oven) are transported to the control system using ductwork and a blower. The ductwork may collect vapors from just one source or from several sources located nearby. The collected vapors are sent to the membrane module. Inside the membrane module, the feed gas is separated into two streams: a concentrated solvent vapor stream ("permeate") and a depleted residue gas stream ("stripped off-gas"). On the permeate side, a vacuum pump pulls a vacuum. An alternate approach is to compress the feed gas with a compressor upstream of the membrane. In either case, an imposed pressure difference across the membrane is the driving force for separation. Most of the organic vapor is drawn through the membrane into the permeate, along with a small amount of air. The stripped off-gas from the membrane is either recycled back to the original vapor source, or may be discharged directly to the atmosphere in some cases. Several options are available for further treatment of the permeate. Possible treatment technologies include:

- Direct condensation (using chilled water/refrigeration; allows solvent recovery, but is viable only at higher solvent concentrations);

- Incineration (with direct flame; this destruction process is usually best for contaminated solvents which are inexpensive to replace [e.g., hydrocarbons] that also have high Btu content); or

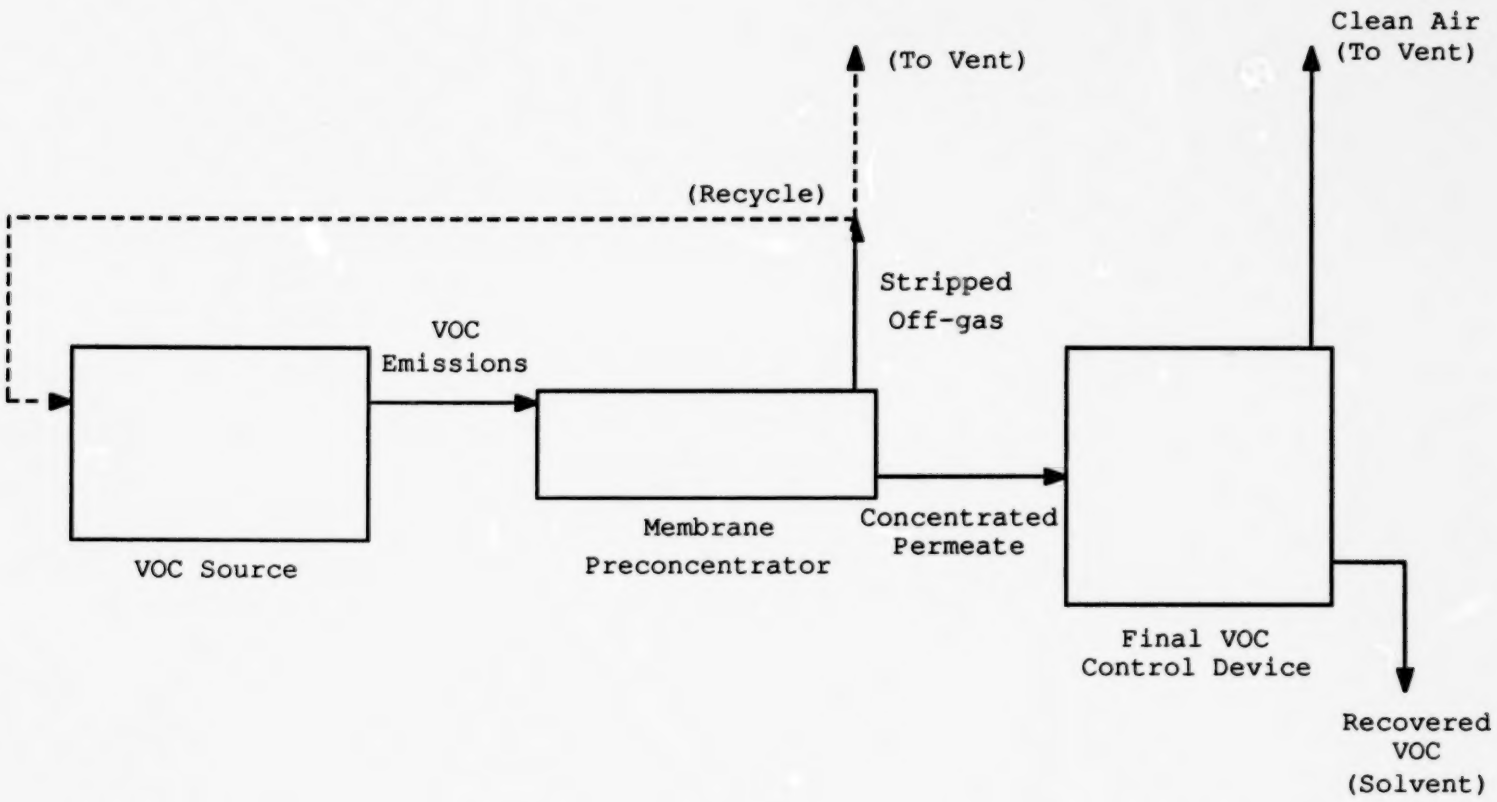

Figure 5-1. Overall Diagram of Membrane Preconcentrator

- Carbon adsorption (in regenerative mode; steam stripping followed by condensation and decanting allows recovery of solvent).

Each of these final control technologies is discussed in more detail in following sections, although the primary focus of the report is on the concept of a membrane preconcentrator in conjunction with carbon adsorption.

SCALE UP

As was explained in Section 4 under "Comparison with Theoretical Model," a material balance model for the single-stage gaseous permeation process was developed by Pan and Habgood (2). In addition to using this model to compare with the laboratory test data, it was also possible to use the model to scale up the test data in order to extrapolate the membrane sizing and performance characteristics for full-scale systems. An example design calculation is presented in Appendix D. Complete design calculations for all cases are listed in Appendix F. The simplifying assumptions are:

- The feed concentration is low (approaching zero), so the simplified version of the equations is valid; and

- An average permeability based on the laboratory test module is a valid approximation of the actual permeability for the full-scale system.

The first assumption was checked by comparing $(1/x_f) > (\alpha^*$ and $1/\gamma)$. The second assumption is not easily checked without a performance test on a larger system. Permeability is a function of the diffusivity (e.g., diffusion coefficient, D) and the solubility (e.g., distribution coefficient, k) for a particular solvent in a given polymer membrane. The laboratory test data was obtained using the membrane material at conditions similar to those which would be employed in full-scale systems. This fact should compensate for the realization that both coefficients D and k increase drastically with an

increase in the initial partial pressure of the solvent. Therefore, it is felt that both assumptions should hold for scale up.

The scale up exercise consisted of designing a multitude of systems at varying flowrates, inlet concentrations, membrane removal efficiencies, membrane selectivities, and for both vacuum pump and compressor based single stage membrane configurations. A complete matrix of design calculations is shown in Table 5-1.

The rationale for the selection of the various parameters is explained below:

1. Inlet flowrates of solvent laden air were varied from 6.2 Nm^3/sec (250 ACFM) to 249 Nm^3/sec (10,000 ACFM) to cover a range of applications that generate airborne solvent emissions;

2. Inlet solvent concentrations of 1000 and 100 ppmv were chosen to cover both typical operations (e.g., solvent degreasers, drying ovens) and also other uses which generate more dilute solvent emissions;

3. Only two solvents (CFC-113 and toluene) were chosen for subsequent system designs and costing because a) they are typical solvents used in vapor degreasing and coating lines, respectively, and b) the permeabilities of these solvents are similar to the other tested solvents. Therefore, the design and cost comparisons are not critically dependent on the solvent selected;

4. Membrane removal efficiencies were varied from 60 to 95% to allow for evaluation of different levels of control. The overall control efficiency of the membrane preconcentrator with a carbon adsorber was slightly lower, due to the 95% control of the carbon adsorber;

TABLE 5-1. DESIGN MATRIX

Inlet flow, Nm^3/sec (ACFM)	Inlet Conc (ppmv)	Solvent	Membrane Removal Efficiencies[a]	Membrane Selectivities	Configuration[b]
6.2 (250)	1000	CFC-113	60,85,95	5,20,200	Vacuum Pump, Compressor
6.2 (250)	100	CFC-113	60,85,95	5,20,200	Vacuum Pump, Compressor
62.2 (2500)	1000	CFC-113	60,85,95	5,20,200	Vacuum Pump, Compressor
62.2 (2500)	100	CFC-113	60,85,95	5,20,200	Vacuum Pump, Compressor
249 (10000)	1000	Toluene	60,85,95	5,20,200	Vacuum Pump, Compressor
249 (10000)	100	Toluene	60,85,95	5,20,200	Vacuum Pump, Compressor

[a]Overall Removal Efficiencies include 95% control by final carbon adsorber:
(60%)(95%)=57% overall efficiency
(85%)(95%)=81% overall efficiency
(95%)(95%)=90% overall efficiency

[b]Configuration is the arrangement of a system with a pressurized feed (i.e., compressor) or a vacuum permeate (i.e., vacuum pump).

5. Membrane selectivities were varied from 5 to 200 to evaluate the effect of membrane thickness on overall system costs. The selected range was chosen to reflect the range of selectivity which has been reported for solvent/N_2 separations with current membranes at typical pressure ratios. Results from the laboratory testing portion of this study showed selectivity values vary between 6 and 7 for most solvents;

6. Two configurations of the membrane system were designed: first, the arrangement using a liquid ring vacuum pump operating at a pressure ratio of 0.10 (Figure 5-2a); second, an alternative arrangement using a turbocompressor (e.g., centrifugal or screw compressor) operating at a pressure ratio of 0.20 (Figure 5-2b). Assuming an initial feed gas pressure of 776 mmHg (15.0 psia), the selected pressure ratios for the two alternative configurations would result in: a suction pressure (at the permeate side) of 78 mmHg (1.5 psia) for the vacuum pump arrangement; and, compression (on the feed side) to 5.1 atm (75 psia) for the compressor arrangement.

To avoid further complicating the comparisons, all the designs were performed on the following common basis:

1. Inlet relative humidity (RH) and temperature were kept at 50% RH at 37.8°C (100°F);

2. Carbon adsorption systems were designed for 95% removal, and all were regenerative systems; and

3. Adsorption isotherms for CFC-113 and toluene on Calgon BPL® carbon were used to design the carbon adsorbers.

4. Overall removal efficiencies for the complete membrane system (i.e., 57 to 90%, including 95% control by the carbon adsorber) are based on a once-through design. If the stripped off-gas was recycled back

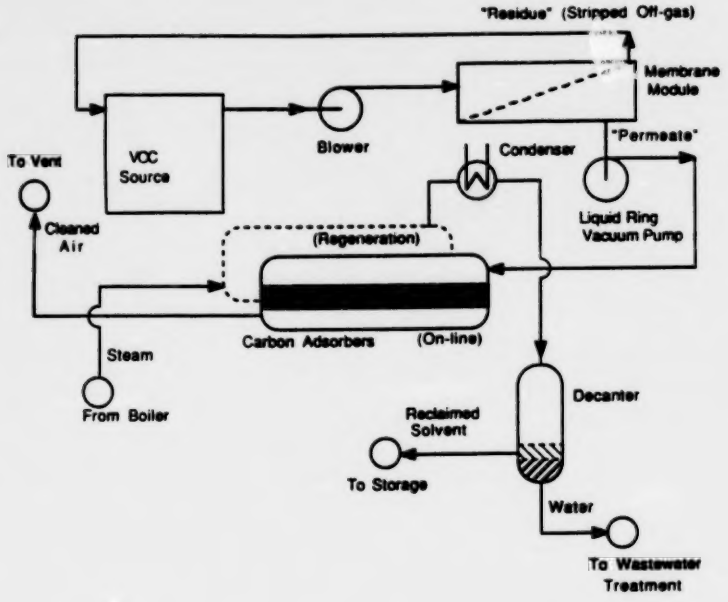

Figure 5-2a. Membrane System with Vacuum Pump

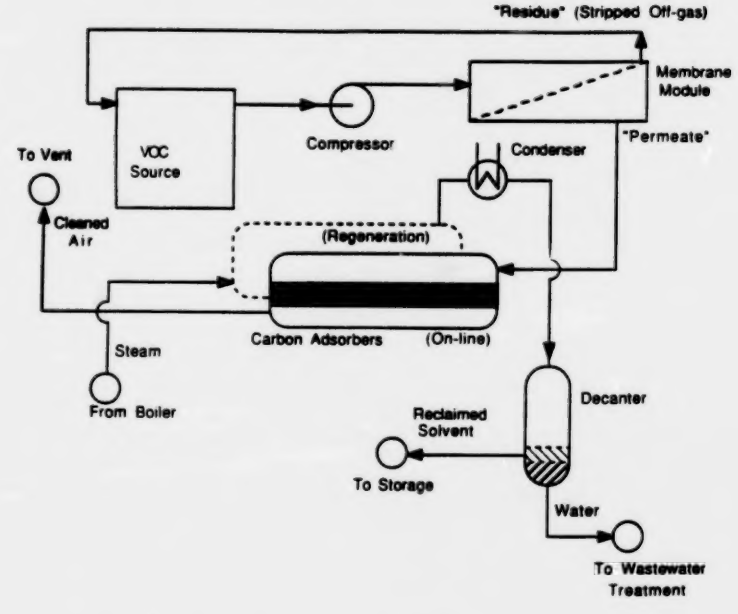

Figure 5-2b. Membrane System with Compressor

to the vapor source, the overall removal efficiency would be higher (approaching 95% control).

INTEGRATION OF OVERALL SYSTEM

This section will discuss several additional factors which may affect the design, costing, and operation of a membrane preconcentrator.

Alternative Arrangements

Besides the arrangements listed above (i.e., single stage membrane with compressor or vacuum pump), other configurations have been developed which offer potential benefits such as greater removal efficiency or higher concentration.

First, adding a purge gas into the permeate side of the membrane can improve the removal efficiency for a given set of conditions by lowering the solvent partial pressure in the permeate. Dilution of the permeate may be more than offset by the increase in solvent flux. However, the amount of purge gas must be carefully calculated to achieve the optimal balance.

Second, it is sometimes desirable to design a multistage membrane system with a recycle stream when higher permeate concentrations are required. These arrangements can involve a multiplicity of recycle flow paths and are often quite complicated. Additional compression or vacuum equipment is usually required. Justification of the added cost and complexity of a multistage recycle system must be shown beforehand (13). MTR has studied and promoted multistage membrane systems as a method to achieve high product recovery.

Third, a membrane configuration known as a "continuous column" has been suggested. This arrangement is claimed to offer a high degree of concentration with less membrane area than other configurations. S. T. Hwang (University of Cincinnati) has published several papers on the subject (14,15,16).

Additional Considerations

There are several other factors which may affect the economics and operation of a membrane preconcentrator. These are discussed below:

Vacuum versus Compression--Both vacuum-based and compression-based membrane systems have advantages and disadvantages. Table 5-2 lists the strengths and weaknesses. At this time, there is insufficient data to conclude which arrangement is better.

Optimum Arrangement of Vacuum System--Preliminary study of the vacuum system hardware which would be required for full-scale systems revealed that several different types of vacuum pumps could be used. Making the correct selection of vacuum pump could have a major impact on the viability of the entire system, since it is the mechanical "heart" of the membrane unit, as well as perhaps the single most expensive component in the system. Listed below are a few of the types of vacuum pumps which may be applicable for a membrane preconcentrator:

- Liquid Ring Pump: Has a wide operating range, both for flow and vacuum level. Only one rotating element, and often uses water as the sealing liquid. The major drawback may be saturation of the permeate gas with water vapor, necessitating the use of a chiller to condense excess water and lower the RH going to the carbon bed.

- Rotary Vane Pump: Limited operating range in flow, so may only be suitable for smaller membrane systems. Rotary vane pumps (also known as Sliding Vane Pumps) offer the advantage of a dry source of vacuum.

- Roots Blower: Wide operating range in flow, but is usually used for higher vacuum levels than a membrane system demands. The Roots blower loses efficiency when a high vacuum is not required, but is also a dry source of vacuum. The Roots blower (also known as a

TABLE 5-2. COMPARISON BETWEEN VACUUM AND COMPRESSION SYSTEMS

System	Strengths	Weaknesses
Vacuum Pump	No aftercooler required. Only the permeate is compressed, so energy is not wasted compressing the entire feed stream. Vacuum operation avoids possibility of condensing solvent inside membrane. Membrane module can be designed for normal pressure operation. Liquid ring pumps are capable of achieving required vacuum levels, efficient, and mechanically simple.	High vacuum levels necessitate large and expensive vacuum pumps. Liquid ring pumps involve separation of entrained liquid from discharge. If water is used as sealing liquid, a chiller may be required to dehumidify prior to carbon adsorption.
Compressor	Slightly less expensive capital cost. Avoid contamination of permeate with additional water; no chiller required. Possible recovery of energy from feed gas compression by passing stripped gas through an expander.	High compression ratios may necessitate addition of aftercooler. Expending energy to compress dilute gas; only small fraction actually permeates. Membrane module might have to be designed as a pressure vessel (ASME-code).

Rotary Lobe Blower) is often combined in a "compound" arrangement with a liquid ring pump.

Effect of Membrane System on Carbon Adsorber--Concentrating the solvent in the permeate offers the potential advantage of reduced gas volume sent to the final control device, in this case, a carbon adsorber. Discussions with vendors of carbon adsorption systems provided some information which must be considered before deciding whether a membrane system is worthwhile.

- If the solvent concentration fed to the carbon adsorber is above 10,000 ppmv or above 25% of the lower explosive limit (LEL), carbon adsorber designers call for dilution air to be added. This would defeat the purpose of obtaining a concentrated solvent feed. Therefore, either the inlet concentration should be designed to be at or below these levels, or the carbon adsorber designer's approval must be obtained.

- Water content in the solvent laden air must be considered, especially when a liquid ring vacuum pump using water as the sealing liquid is part of the system. As mentioned before, entrained water droplets from the liquid ring pump are removed, but the exiting gas is nonetheless saturated with water vapor. Without a chiller to condense and remove the excess water, the adsorber beds would have to be sized for additional carbon. The unknown effect of solvent removal in the chiller condenser would also complicate the design.

Viability of Other Final Control Technologies--As was mentioned earlier in this section, both direct condensation and incineration are possible alternatives to carbon adsorption as final control technologies. However, they are not without their own limitations as well.

For instance, direct condensation using chilled water or other low temperature refrigerants is a possible alternative. This type of system has been patented by MTR (7) and discussed in several papers. Commercial systems are currently in use for recovery of gasoline vapors from bulk storage

terminals. The best applications for this type of system are those with high solvent concentrations from the source (i.e., 5-10,000 ppmv and higher). However, its applicability to dilute solvent vapor streams (i.e., less than 1000 ppmv) would require either highly selective membranes (which would in turn require larger areas), multiple stages of membranes with recycle (which would require additional vacuum or compression equipment), or very low temperatures (with high power requirements).

Likewise, incineration in a direct flame is a possible alternative, especially if the solvent vapor is flammable, contaminated, has a high Btu value, and is inexpensive to replace. Combination systems using carbon adsorption and incineration (such as Calgon Carbon's CADRE® system) are commercially available. Disadvantages to incineration for use with a membrane system include:

- Difficulty in handling chlorinated or fluorinated solvents (i.e., corrosive products of combustion such as HCl and HF);
- Low concentrations or nonflammable solvents would require supplemental fuel;
- Unlikely that small incinerators would be easily permitted or accepted by users; and
- Loss of a recoverable product.

SECTION 6

COST ALGORITHMS

VENDOR SURVEY AND LITERATURE DATA

Preparation of the cost estimates for the systems previously designed (as discussed in Section 5) involved first obtaining baseline cost data for all major capital components. This cost data was obtained through written Requests for Quotations from equipment vendors, telephone contacts with equipment vendors, and available literature data.

Requests for Quotes

In order to obtain current cost data for vacuum pumps and carbon adsorbers (two of the most expensive elements in the overall system), vendors were contacted and requested to provide cost quotes for selected equipment. The following vendors were contacted and sent letters requesting budget cost quotes:

1. Carbon Adsorbers -
   American Ceca Corp. (AMCEC)
   RaySolv Inc.
   Barnebey and Sutcliffe Corp.

2. Vacuum Pumps -
   Nash Engineering Co.
   SIHI Pumps, Inc.
   Ochsner Pumps
   Intervac Corp.
   Edwards High Vacuum

Leybold Heraeus Vacuum Products
Balzers
Kinney Vacuum Co.

Unfortunately, this approach was generally unsuccessful in providing any cost data. Typically, the vendor simply declined to quote. Also, some of the vacuum pump vendors could not supply pumps capable of handling the high flowrates.

Telephone Contacts

Because of the poor response to written requests for cost data, additional telephone contacts were made to follow-up on the letters or to establish new contacts with other vendors. The following phone contacts were made to obtain additional cost data:

1. Carbon Adsorbers -
   Mr. Bob Spencer (Allied Signal/Baron Blakeslee)
   Mr. Tom Cannon (VIC Manufacturing)

2. Vacuum Pumps -
   Mr. Tom Walker (SIHI Pumps, Inc.)
   Mr. Mike Whiteside (Telesis High Vacuum for Kinney Vacuum)
   Mr. Lou Sleigher (Balzers)

Literature Sources

Available literature data was used for most of the cost estimates. These estimates were cross-checked with the quoted prices from vendor contacts. The sources of literature cost data are listed in Table 6-1.

Summary of Cost Data

Because of the diversity of cost data (some from vendor quotes, some from telephone contacts, and other data from literature sources), this section will

Leybold Heraeus Vacuum Products
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Because of the diversity of cost data (some from vendor quotes, some from telephone contacts, and other data from literature sources), this section will

TABLE 6-1. LITERATURE COST DATA

1. Membrane Modules -

   Development of Synthetic Membranes for Gas and Vapor Separation. Strathman et al. Pure and Applied Chemistry. Vol. 58, No. 12. 1986 (Ref. 8).

   The Separation of Organic Vapors from Air. Peinemann et al. AIChE Symp. Ser. 250(82):19, 1986 (Ref. 3).

2. Carbon Adsorbers -

   Capital and Operating Costs of Selected Air Pollution Control Systems - I. R.B. Neveril et al. Journal of the Air Pollution Control Association. Vol. 28, No. 8. August 1978 (Ref. 10).

   The Cost of Controlling Organic Emissions. Kittleman and Akell. Chemical Engineering Progress. April 1978 (Ref. 11).

3. Vacuum Pumps -

   Chemical Engineering (Dec. 14, 1981) "Selecting Vacuum Systems," by J.L. Ryans and S. Croll (Ref. 9).

4. Compressors -

   Plant Design and Economics for Chemical Engineers (3rd Ed.). Peters and Timmerhaus. 1980 (Ref. 12).

present a discussion of the various cost values and provide a rationale for the foundation of the cost analysis which follows.

Tables 6-2-A through 6-2-D present a listing of the available cost data for the four major cost items (i.e., membrane, carbon adsorber, vacuum pump, and compressor), along with the source of the data. Some cost data may not list a source if it was requested to be kept confidential.

It is important to note a few points about the capital cost estimates for carbon adsorbers. First, it was difficult to obtain good cost data for regenerative carbon adsorbers at the low flowrate range [i.e., less than 28 Nm^3/sec (1000 SCFM)]. Only a few data points were available, and it was found that these small units were nearly as expensive as their larger counterparts. Discussions with carbon adsorber vendors revealed that although reduced volume flow can allow slightly smaller components, the fabrication and materials costs are nearly the same. Likewise, the carbon requirements and operating costs for both membrane-concentrated permeate vapor and direct untreated flow from the source were almost equal, since carbon requirements are governed mostly by the mass of solvent to be adsorbed, an amount that is nearly the same for both cases (the higher concentration of solvent from the membrane unit does provide a greater driving force and hence slightly less carbon). Presumably, at very low flows, a non-regenerative "canister" type of unit might be more attractive.

RESULTS OF COST ANALYSIS

Capital Cost Comparison

Using the system designs for the CFC-113 and toluene systems outlined previously in Section 5, capital costs were estimated for complete systems using a membrane preconcentrator. These costs were also compared to the capital cost for a carbon adsorber alone handling the same duty. Figures 6-1 through 6-6 present the total installed capital cost for these cases. Table 6-3 provides an example cost comparison showing the effect of the membrane on

TABLE 6-2-A. LISTING OF CAPITAL COST DATA AND SELECTION OF COST BASIS

MEMBRANE COST DATA			
	Cost Data		Source
A.	Membrane Module: Other System Costs:	$\$200/m^2$ $\$200/m^2$	Nitto Denko (Japan)
B.	Membrane Module: Other System Costs:	$\$40/m^2$ $\$40/m^2$	Reference 3
C.	Membrane Module: Other System Costs:	$\$150/m^2$ $\$188/m^2$	Reference 8

Cost Basis Selected for This Report:

Membrane Module[a]:	$\$100/m^2$
Other System Costs[b]:	$\$50/m^2$

[a]Cost basis for membrane module based on an approximate average of cost data: $(200 + 400 + 150)/3 \approx \$130/m^2$ rounded to $\$100/m^2$

[b]Cost basis for other system costs was reduced to $\$50/m^2$ since cost data from other sources included either items costed separately in this work (e.g., vacuum pump), or not applicable (e.g., condenser-chiller).

TABLE 6-2-B. LISTING OF CAPITAL COST DATA AND SELECTION OF COST BASIS

CARBON ADSORBER COST DATA

Inlet Solvent-Laden Air Flowrate (SCFM)	Carbon Capacity (Pounds per Adsorber)	Number of Adsorbers	Material of Construction[a]	Capital Cost (P=purchased equipment; I=installed equipment)	Source
250		2		$45,000 (P)	Vendor A[b]
600	500	2	304SS	$74,500 (P)	"
1,000	900	2	304SS	$84,500 (P)	"
1,000	900	2	316SS	$115,600 (P)	"
1,600	1,400	2	304SS	$96,700 (P)	"
300		2	304SS	$56,000 (P)	Vendor B[c]
1,000		2	304SS	$62,500 (P)	"
2,000		2	304SS	$125,000 (P)	"
2,000		2	304SS	$200,000 (I)	"
8,000		3	304SS	$200,000 (P)	"
300		2	316SS	$164,500 (I)	Vendor C[d]
1,000		2	316SS	$179,400 (I)	"
2,000		2	316SS	$227,300 (I)	"
4,000		3	316SS	$296,400 (I)	"
20,000		3	316SS	$583,300 (I)	"

(Continued)

TABLE 6-2-B. (Continued)

Cost Basis Selected for This Report:

Vendor capital cost algorithm was developed and used for the carbon adsorber system capital costs, with the following exceptions:

- A single cost estimate of $45,000 was used for adsorber vessels for "carbon adsorption only" at an inlet flowrate of 250 ACFM. This was done to correct for the insensitivity of the original cost algorithm at low flowrates. That is, it appears that the original cost algorithm overestimated the installed costs at the lowest flowrate.
- The single estimate was also used for all comparisons where the inlet flowrate was 250 ACFM or less (i.e., with and without the membrane preconcentrator). Therefore, even when the membrane reduced the inlet flow to the carbon adsorber, the adsorber cost was kept constant. This is due to the fact that below 250 ACFM, costs of material and fabrication may remain nearly constant. It may be argued that we have simply shifted the point of constant cost to lower flowrates.
- For the cases with inlet flowrate of 2,500 ACFM or higher, the original cost algorithm was used for all comparisons (i.e., both with and without the membrane preconcentrator).

[a]Appropriate materials of construction must be selected based on the type of solvent being handled. Resistant base metals such as 316 SS are often used with corrosive materials, and the costs were based on using 316 SS. Capital costs of carbon adsorbers will vary considerably depending on the materials of construction. For stable aromatics (such as toluene), mild steel with a coating may be used. At the other extreme, unstabilized halogenated solvents may form acidic hydrolysis products during steam regeneration. Chloride stress cracking may prohibit using ferritic stainless steels, necessitating exotic alloys such as Hastelloy or Monel.

[b]Vendor A quoted standard carbon adsorbor package which is prefabricated with a fully automatic control panel, safety interlocks, inlet air filter, fan, condenser, and decanter. Additional cost of $12,000 for breakthrough analyzer and recorder to control regeneration cycles (included on fourth unit listed).

[c]Vendor B quoted carbon adsorber packages (which are built to order) and include a 25% adder for skid mounting on the first two units.

[d]Vendor C installed costs of several regenerative carbon adsorption systems based on vendor cost quotes. All systems were designed for 95% removal of a 2,000 ppm methylene chloride-in-air inlet gas stream.

TABLE 6-2-C. LISTING OF CAPITAL COST DATA AND SELECTION OF COST BASIS

VACUUM PUMP COST DATA

	Capacity (ACFM @ 27 in Hg vacuum)	Driver HP	Installed Cost	Source
A.	3,000	200	$99,000[a]	Vendor C[b]
B.	600		$55,500[a]	Vendor D[c]
	6,000		$559,000[a]	"
	23,400		$1,287,000[a]	"
C.	500	30	$58,000	Ref. 9[b,d]
	5,000	300	$190,000	"
	21,000		$370,000	"

Cost Basis Selected for This Report:

Design/cost equation in Ref. 9 with escalation factor of 1.232.

[a]Using installation cost = 150% of purchased equipment cost.
[b]Liquid ring pump.
[c]Rotary (roots) blower.
[d]Using design/cost equation in Ref. 9, escalated to 1st Qtr. '89 (C.E. Plant Cost Index for Pumps and Comprs: 473/384 = 1.232).

TABLE 6-2-D. LISTING OF CAPITAL COST DATA AND SELECTION OF COST BASIS

COMPRESSOR COST DATA

Capital Cost[a] (1989 dollars) = 222.8 [(capacity in $ft^3/min)^{0.903}$]

Capital Cost[b] (1989 dollars) = 2272 [(capacity in $ft^3/min)^{.516}$]

Source: Reference 13 (Figures 13-46 and 13-52, respectively), escalated to 1st Qtr. '89 (C.E. Plant Cost Index: 351/230 = 1.526).

Cost Basis Selected for This Report:

A. For 250 ACFM cases: Used Figures 13-46 from Reference 12 with escalation factor of 1.526.

B. For 2,500 and 10,000 ACFM cases: Used Figures 13-52 from Reference 12 with escalation factor of 1.526.

[a]For helical screw compressors at 150 psi discharge from 130 cfm - 800 cfm.

[b]For turboblowers at 30 psi discharge from 1800 cfm - 16,000 cfm.

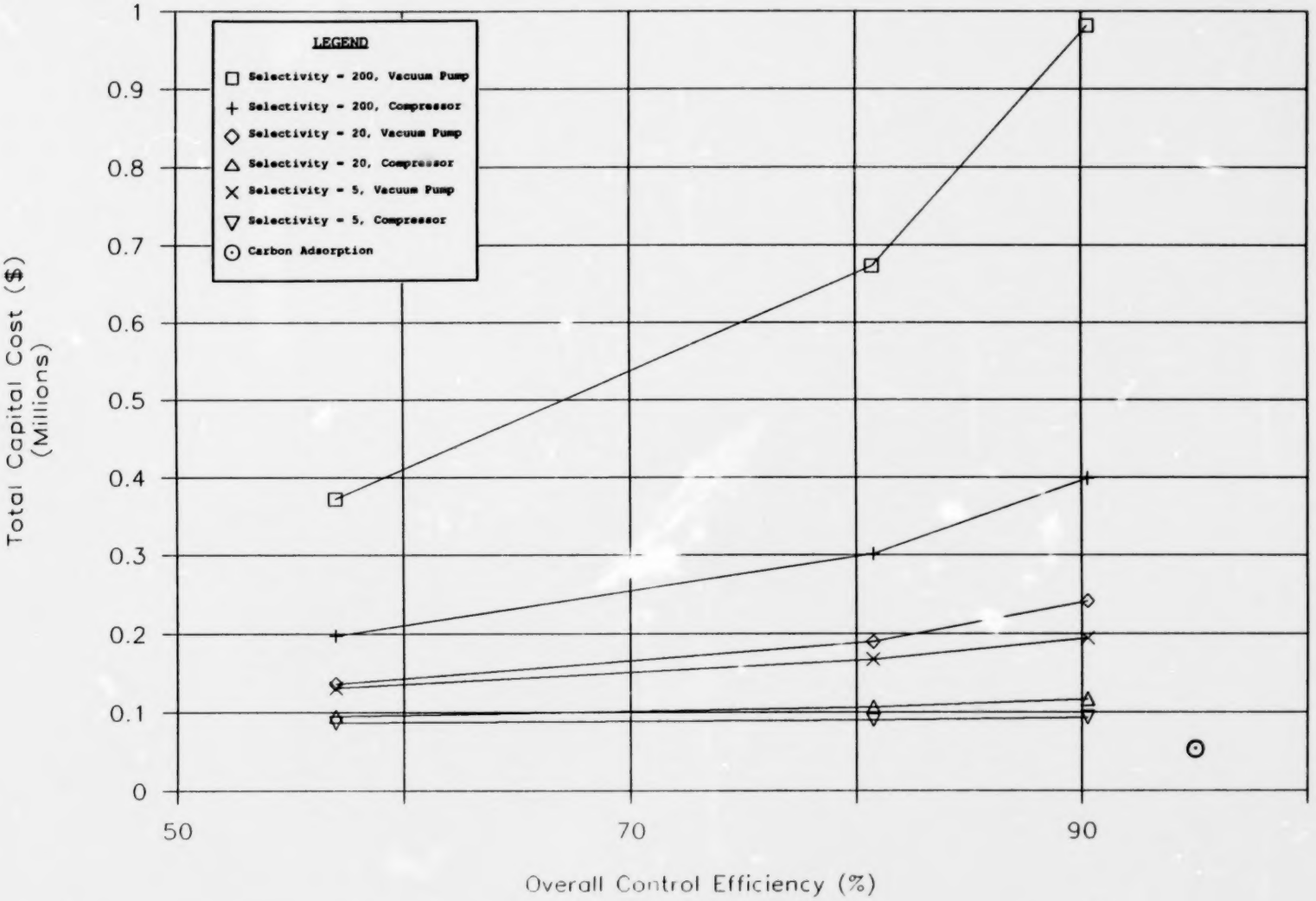

Figure 6-1. Capital Cost Comparison (250 ACFM, 1000 ppm CFC-113 feed)

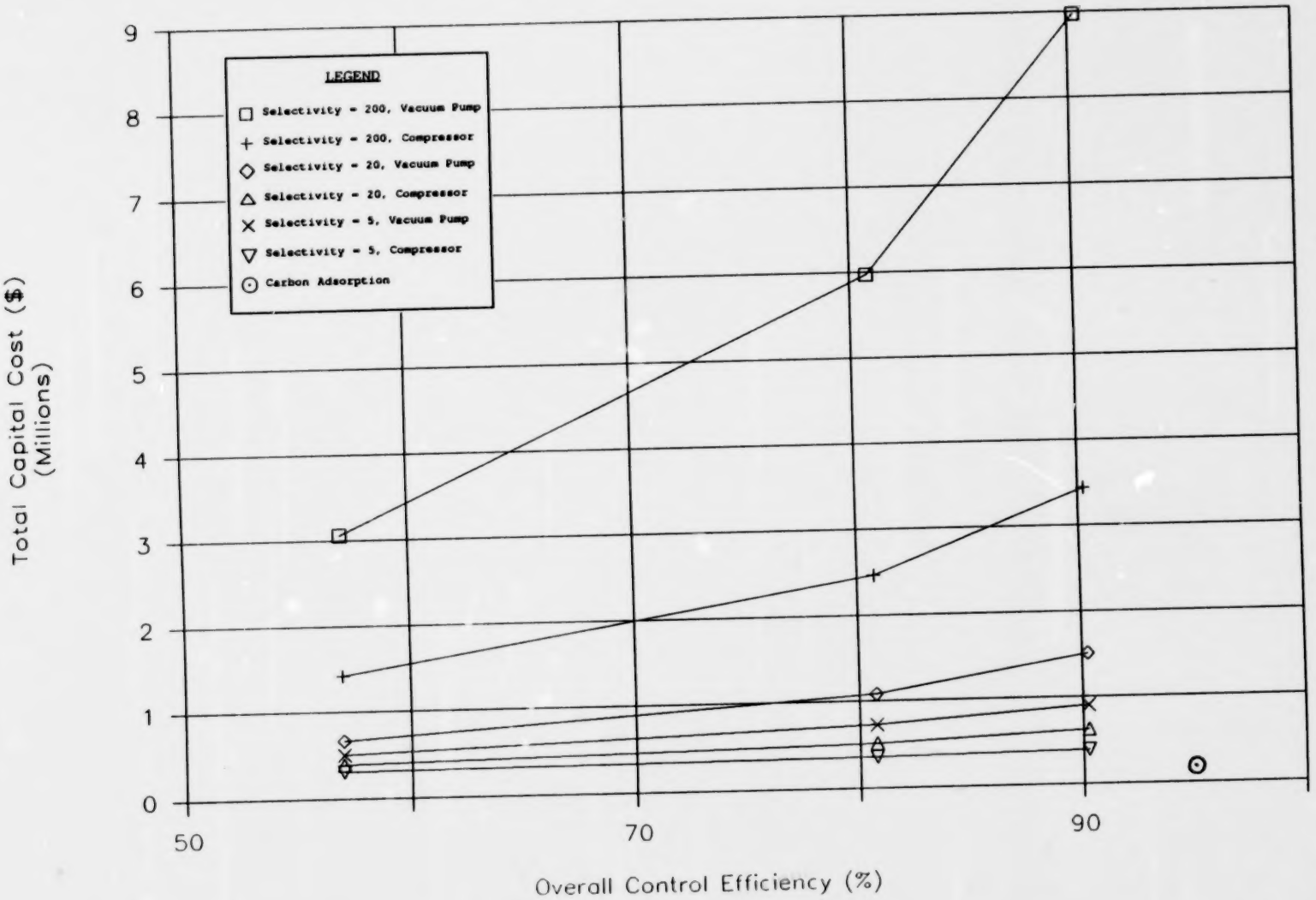

Figure 6-2. Capital Cost Comparison (2500 ACFM, 1000 ppm CFC-113 feed)

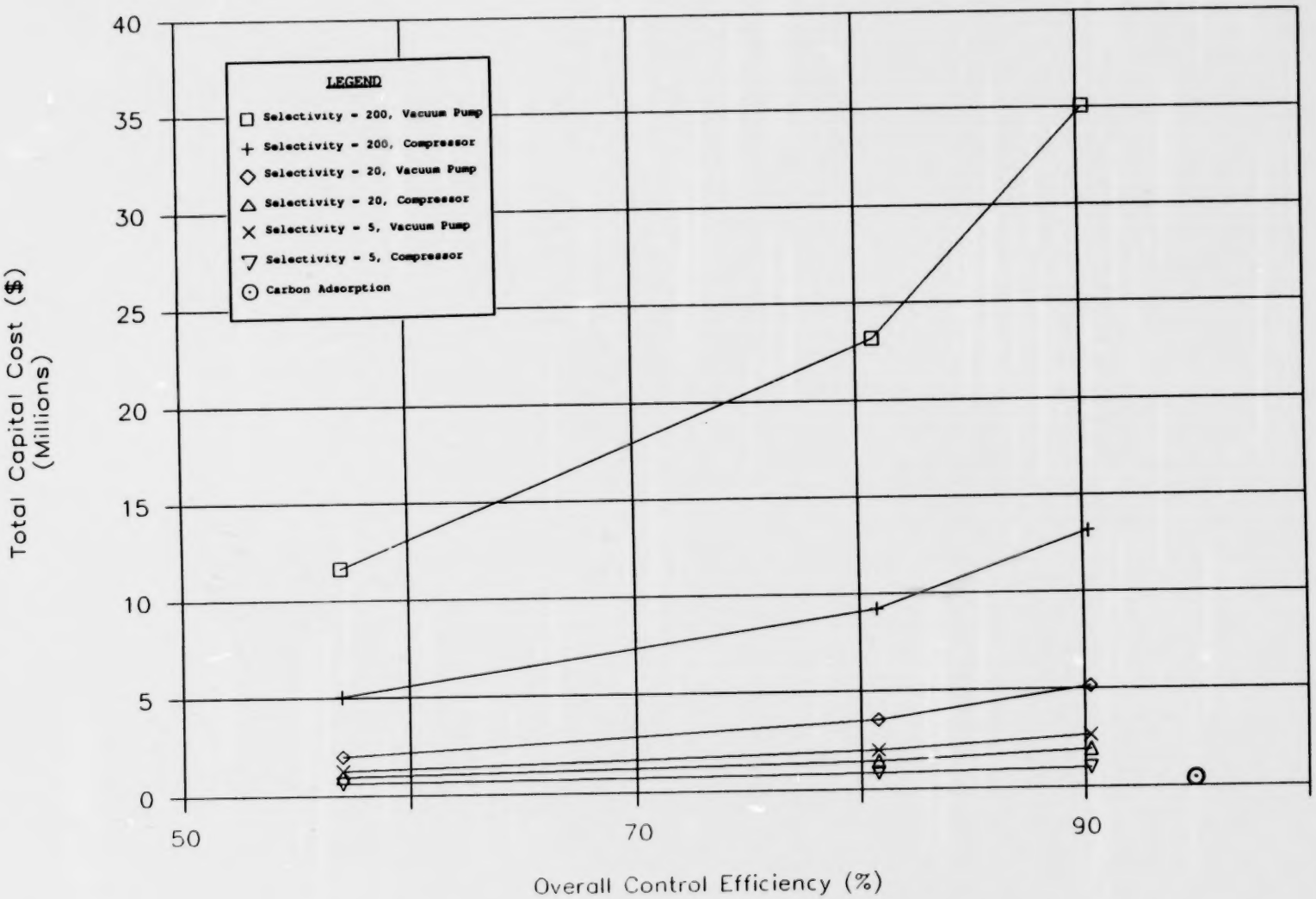

Figure 6-3. Capital Cost Comparison (10000 ACFM, 1000 ppm Toluene feed)

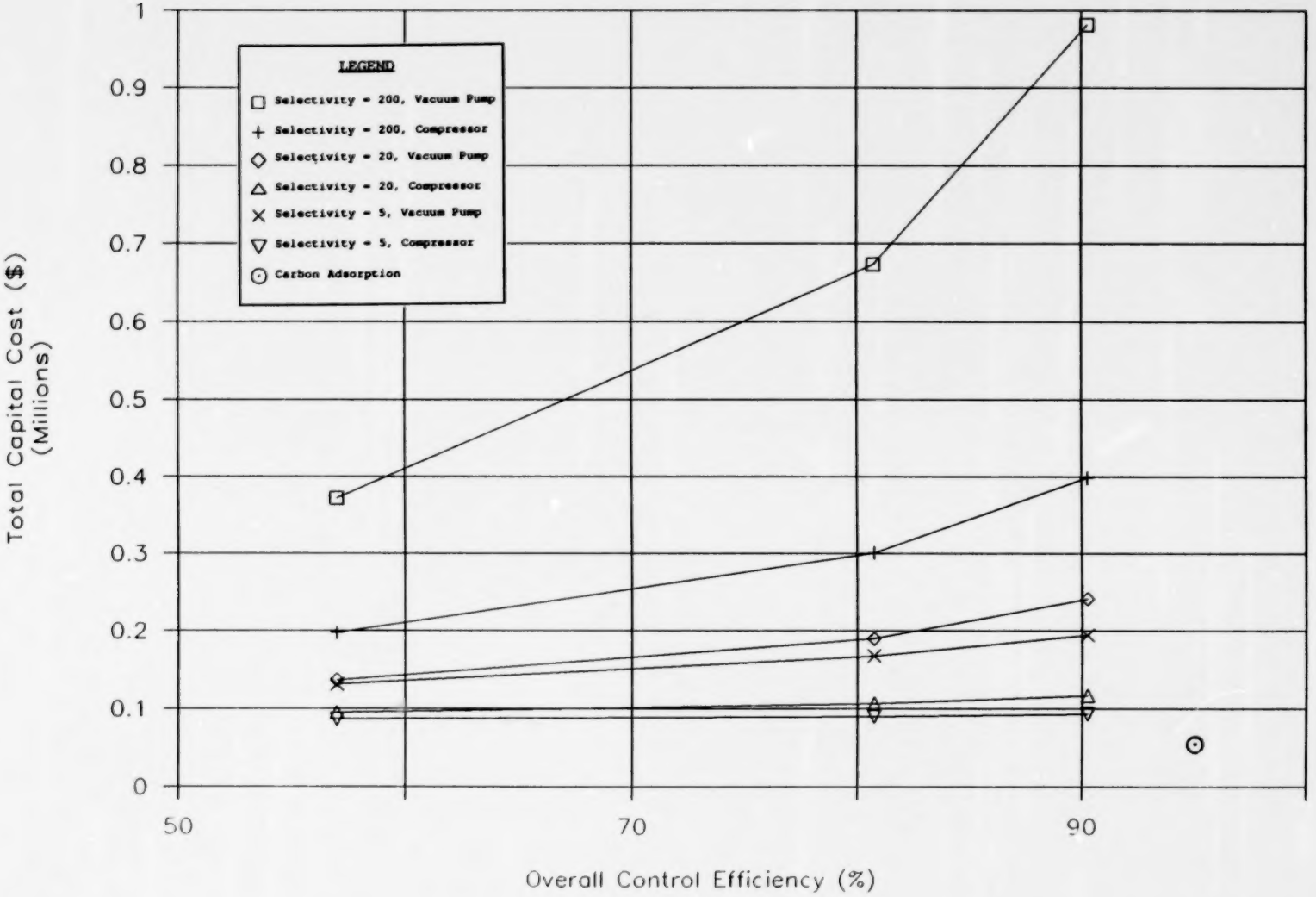

Figure 6-4. Capital Cost Comparison (250 ACFM, 100 ppm CFC-113 feed)

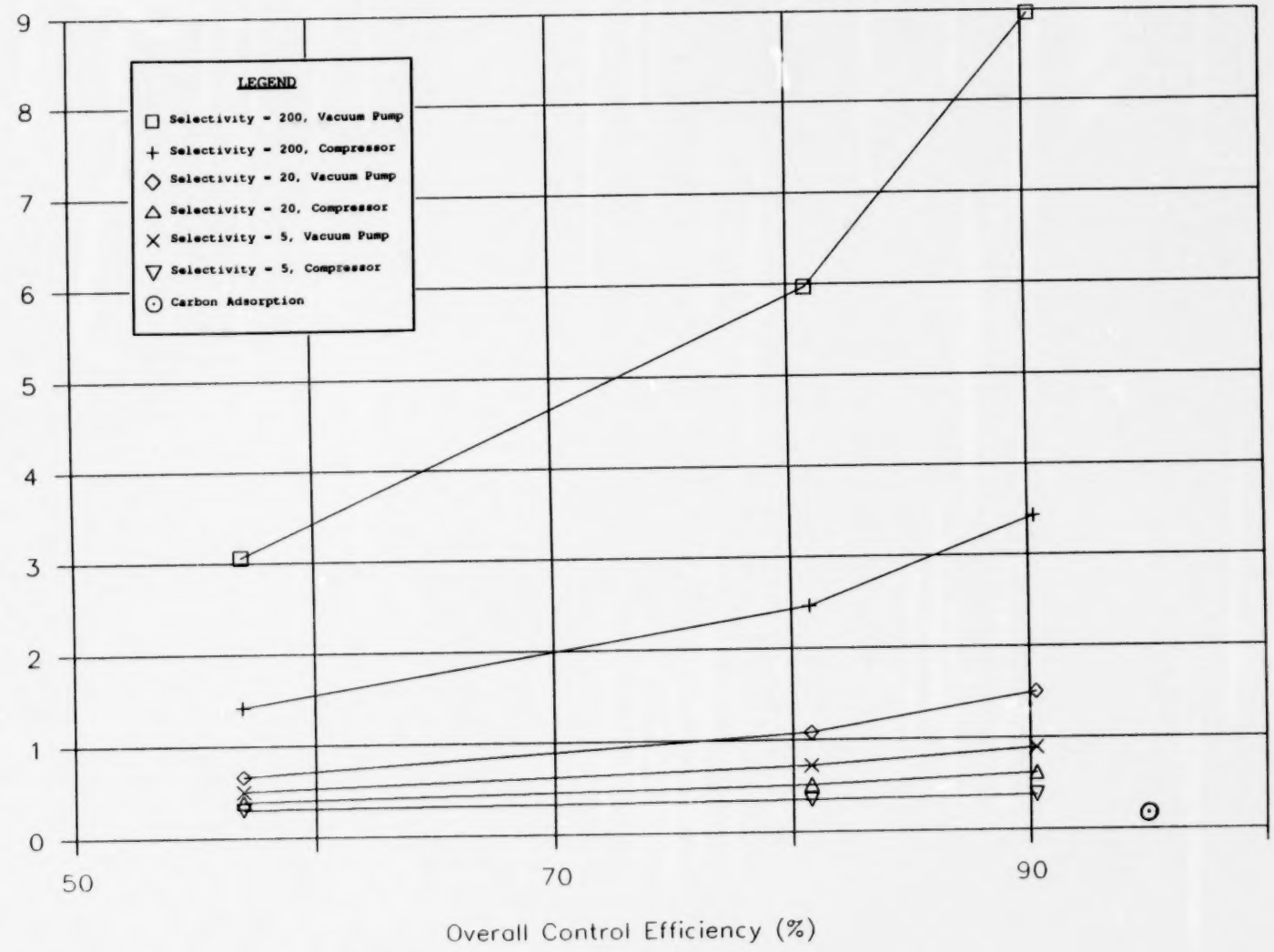

Figure 6-5. Capital Cost Comparison (2500 ACFM, 100 ppm CFC-113 feed)

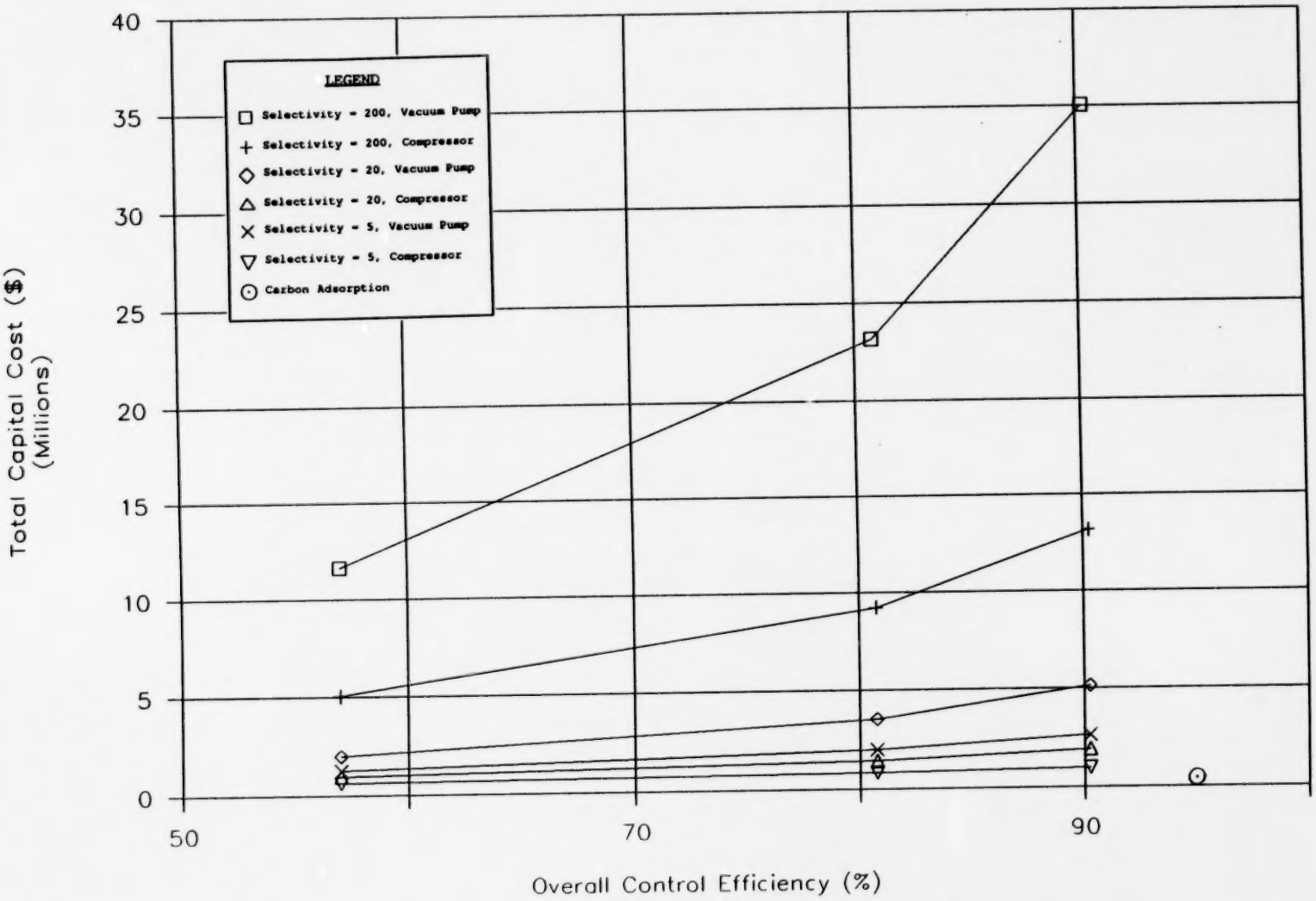

Figure 6-6. Capital Cost Comparison (10000 ACFM, 100 ppm Toluene feed)

TABLE 6-3. EXAMPLE CAPITAL COST COMPARISON[a]

System Flow Rate, Nm^3/s (cfm) = 62.2 (2500 ACFM) Inlet Solvent Concentration = 1000 ppm CFC-113	A. Carbon Adsorber Only[b]	B. Membrane Preconcent. w/Vac Pump[c]	C. Membrane Preconcent. w/Compr[d]
(Costs in 1st qtr-1989 dollars)			
Installed Equipment Costs[e]			
Membrane Module	Not Applic.	$259,900	$89,600
Auxiliary Equipment (Interconnecting piping, controls, etc.)	Not Applic.	$129,900	$44,800
Vacuum Pump (includes motor driver)	Not Applic.	$146,100	Not Applic.
Compressor (includes motor driver)	Not Applic.	Not Applic.	$128,000
Carbon Adsorption System			
- Adsorber Vessels	$152,500	$124,700	$127,000
- Duct work	$5,800	$4,600	$4,500
- Fans	$2,600	$2,000	$2,000
- Carbon	$3,200	$800	$900
Total Capital Costs	$164,100	$668,000	$396,800

Notes:

[a] Indirect costs (e.g., Engineering & Supervision, Construction Expenses, Contractor Fees, and other miscellaneous charges) are assumed to be included in the installation cost. These prefabricated, skid-mounted solvent recovery systems should not have high costs of field erection and start-up.

[b] Achieves 95 percent removal efficiency.

[c] Achieves 57 percent removal efficiency (overall).

[d] Achieves 57 percent removal efficiency (overall).

[e] Installed costs from literature sources, or estimated as 150 percent of purchased equipment cost.

capital costs versus direct carbon adsorption. A sample calculation is shown in Appendix D, and a listing of all cost calculations is presented in Appendix G.

It is evident from these graphs that the membrane system was more expensive in all cases. The increased capital cost ranged from roughly twice (2 times) as costly to over one hundred (100) times as costly, depending on the cases under consideration. The underlying reason for the higher costs is that although the membrane unit is able to reduce the volume flow of solvent vapor to the carbon adsorber, and thereby allow for a smaller carbon adsorber, this reduction is not sufficient to provide cost savings which cover the added expense associated with the membrane and vacuum pump or compressor.

Operating (Annual) Cost Comparison

In an analogous fashion to the capital costs described above, the annual costs were compared for systems with and without a membrane preconcentrator. In order to compare the costs on a common basis in terms of the amount of solvent controlled, this section presents an annualized cost effectiveness result. Table 6-4 presents a listing of the unit costs of various charges for operating labor, utilities, and interest charges. Figures 6-7 to 6-12 present the comparisons of annualized cost for the membrane systems versus carbon adsorption alone. Table 6-5 presents an example cost comparison showing the effect of the membrane on annualized cost effectiveness. A sample calculation of annualized costs is shown in Appendix D, and a complete listing of all annualized cost calculations is presented in Appendix G.

The annualized costs for the membrane augmented system are uniformly higher than carbon adsorption alone. The membrane system costs ranged from about twice (2 times) as expensive to over one hundred thirty (130) times as expensive as straight carbon adsorption for the cases examined. As was seen earlier in the capital cost comparison, although the membrane system allowed slightly lower operating costs for the downsized carbon adsorber, the pumping (or compression) costs, membrane replacement costs and higher capital recovery costs outweighed these savings.

TABLE 6-4. LISTING OF UNIT COSTS IN ANNUALIZED COST COMPARISON

Annualized Cost Element	Unit Cost	Unit Cost
Direct Cost Elements		
Operating Labor[a]	$28.61 /day	
Maintenance Labor[a]	$19.07 /hr	
Steam[b]	$0.0121 /kg	($0.0055 /lb)
Electricity	$0.0572 /kWh	
Cooling Water	$0.008 /$m^3$	($0.03 /1000 gal)
Wastewater Disposal[c]	$0.018 /kg	($0.008 /lb)
Replacement Carbon[d]	$4.69 /kg	($2.13 /lb)
Replacement Membranes[d]	$100 /$m^2$	($9.30 /$ft^2$)
Indirect Cost Elements		
Capital Recovery Factor[e]	0.1627	
Credits		
Recovered Solvent Credit[f]	$0.1375 /kg	($0.0625 /lb)

Notes:

[a] Operating and maintenance labor were assumed to be the same for all units.

[b] Cost for generating steam in existing boiler.

[c] Cost for disposal without an air stripper.

[d] Three (3) year life for carbon and membranes.

[e] Applied to total capital investment for 10 years at 10 percent prevailing interest.

[f] Based on assumption that recovered solvent is worth 50 percent of new solvent (new solvent cost assumed to be $0.264/L ($1.00/gal)). Although a full credit of new solvent cost is sometimes allowed towards the reclaimed solvent, the 50% credit chosen here assumes that the solvent from the carbon adsorber will require some additional treatment. Examples of post-recovery treatments include: addition of stabilizers which are lost during regneration, and dehydration to remove residual water.

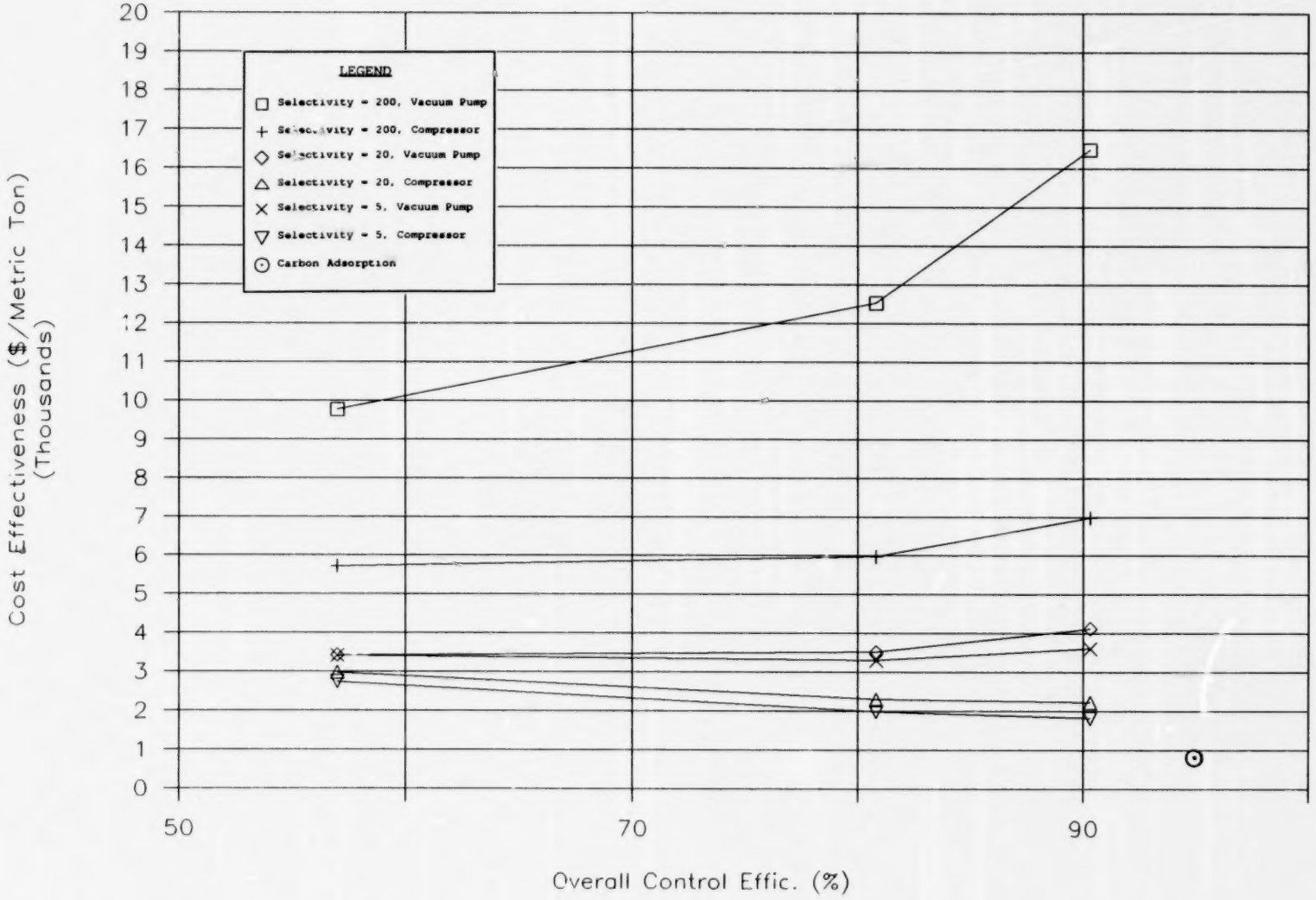

Figure 6-7. Annualized Cost Comparison (250 ACFM, 1000 ppm CFC-113 feed)

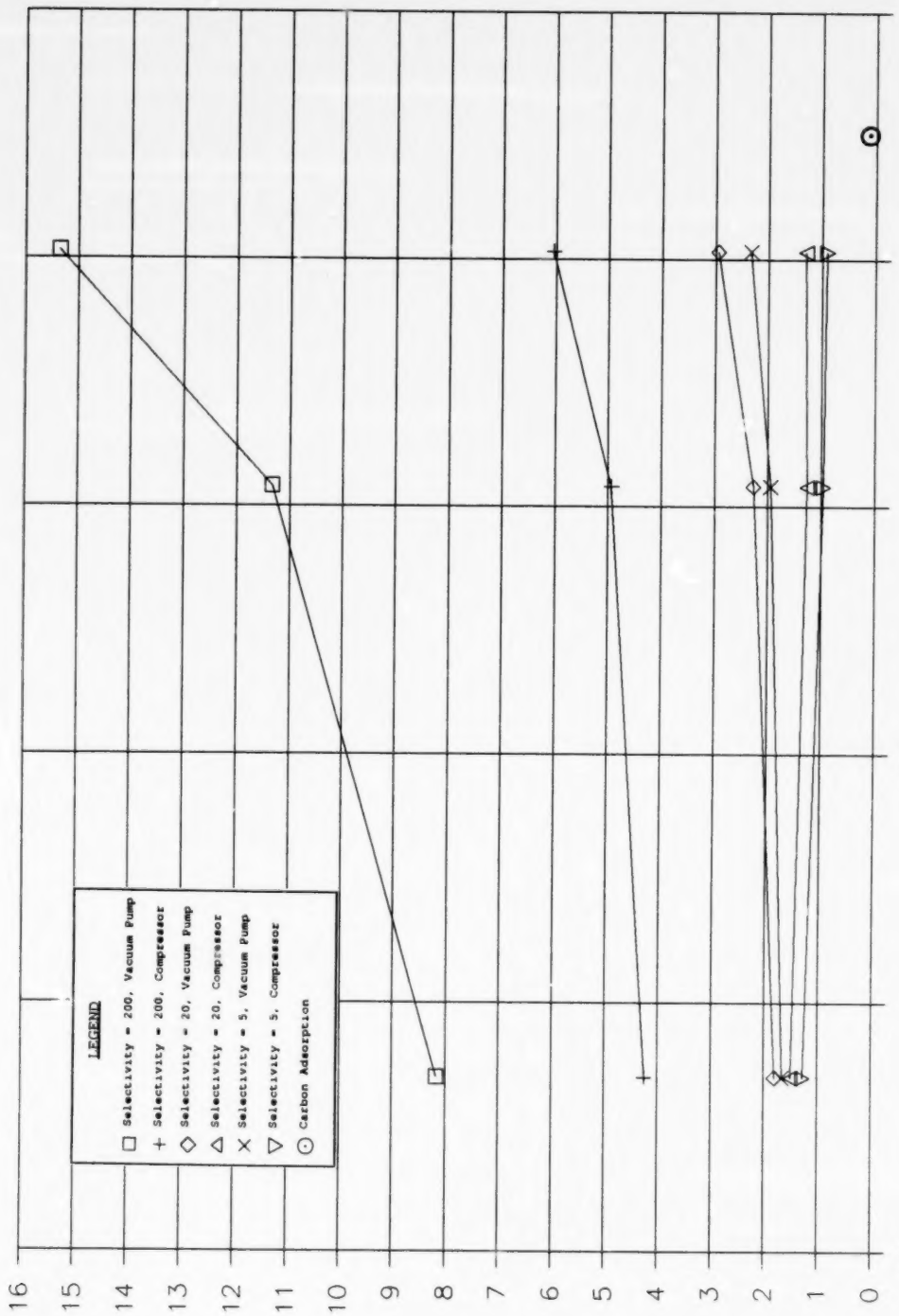

Figure 6-8. Annualized Cost Comparison (2500 ACFM, 1000 ppm CFC-113 feed)

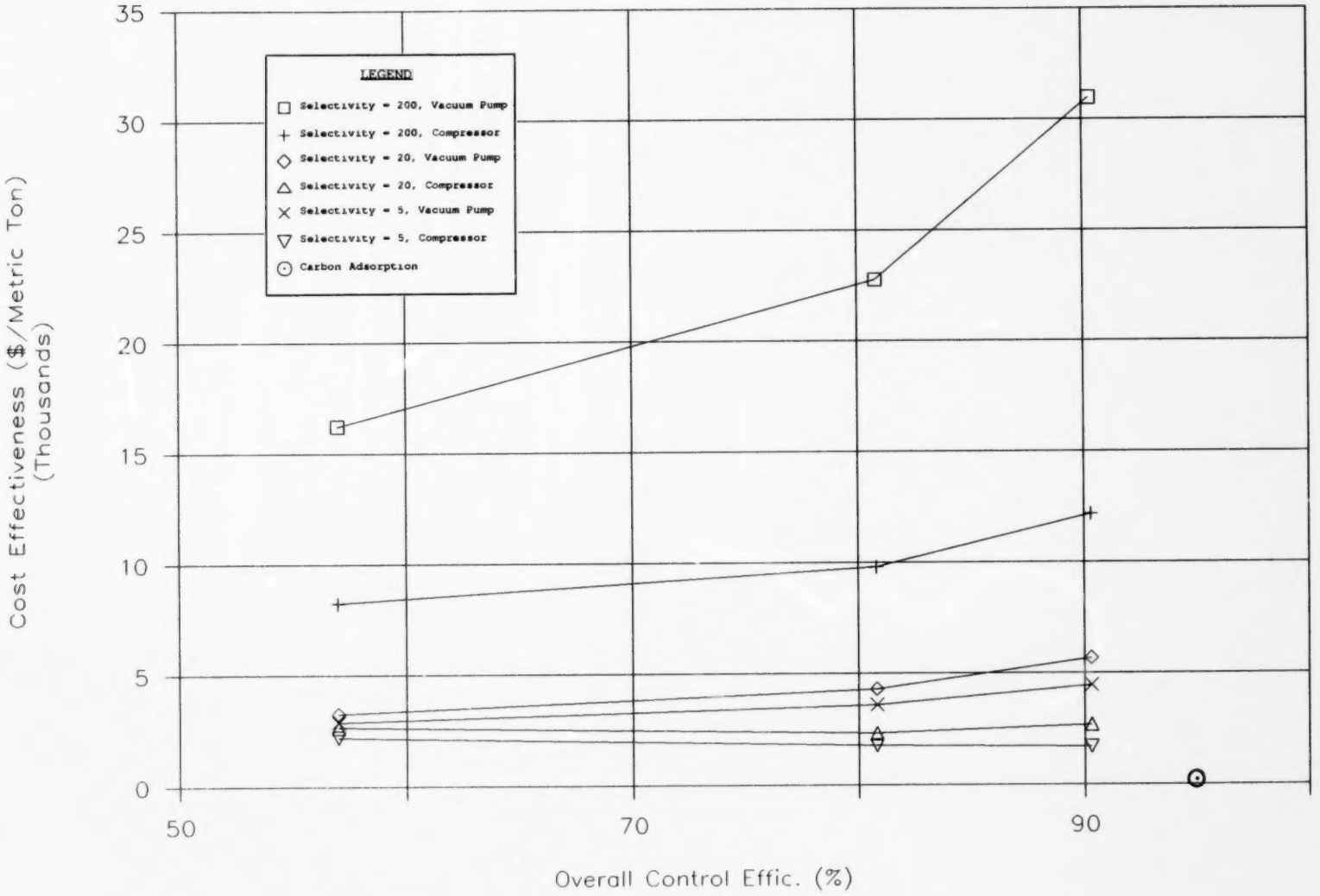

Figure 6-9. Annualized Cost Comparison (10000 ACFM, 1000 ppm Toluene feed)

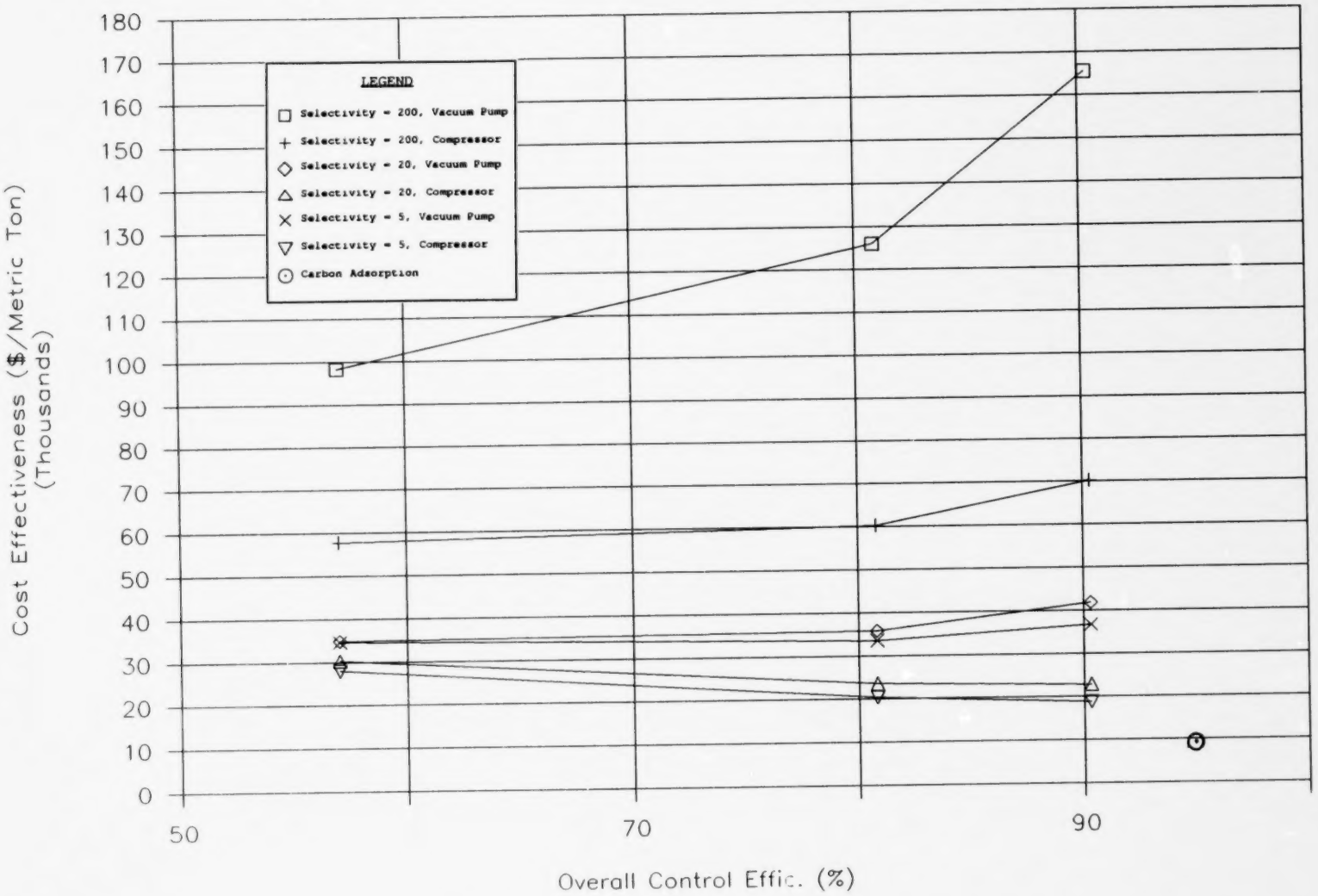

Figure 6-10. Annualized Cost Comparison (250 ACFM, 100 ppm CFC-113 feed)

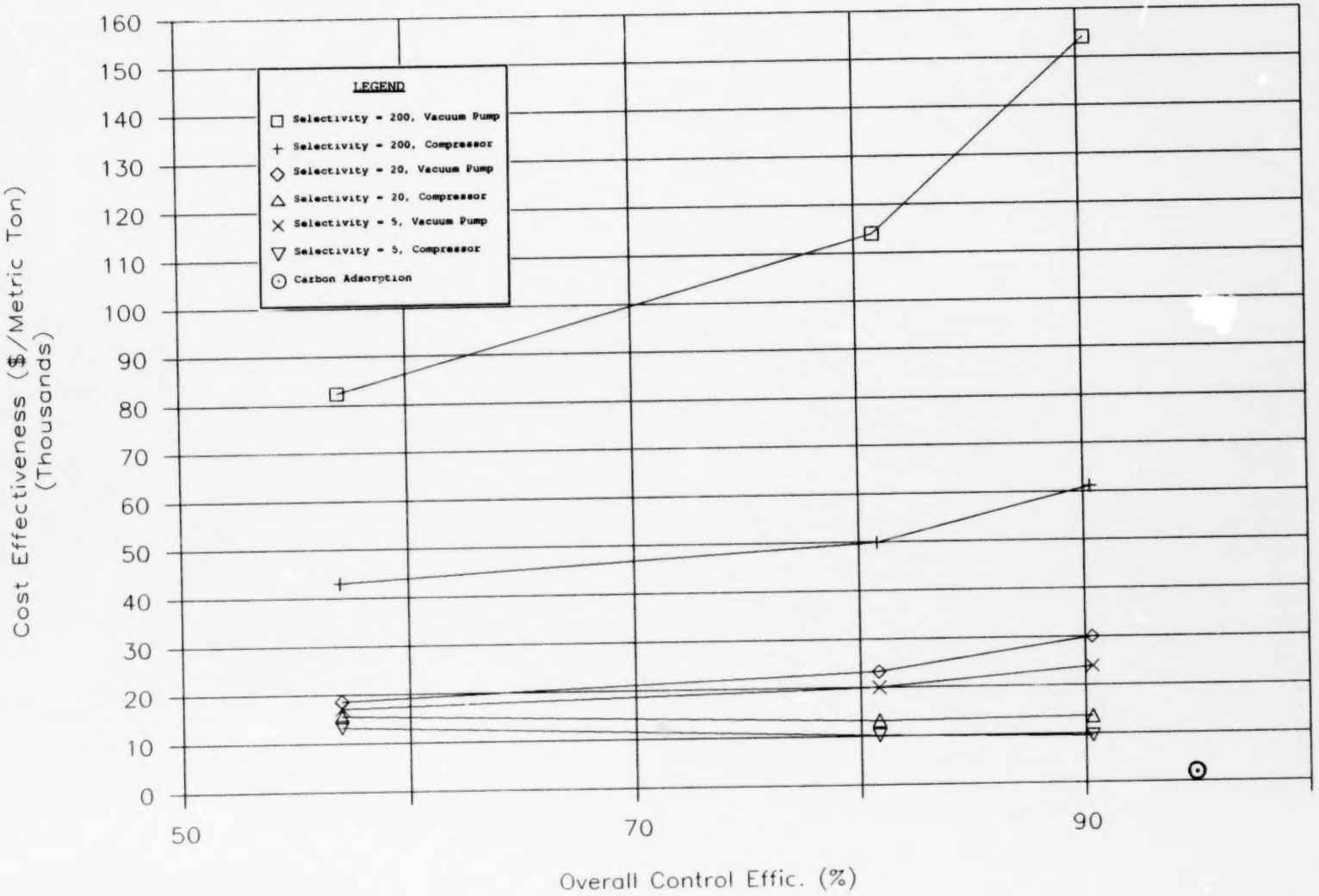

Figure 6-11. Annualized Cost Comparison (2500 ACFM, 100 ppm CFC-113 feed)

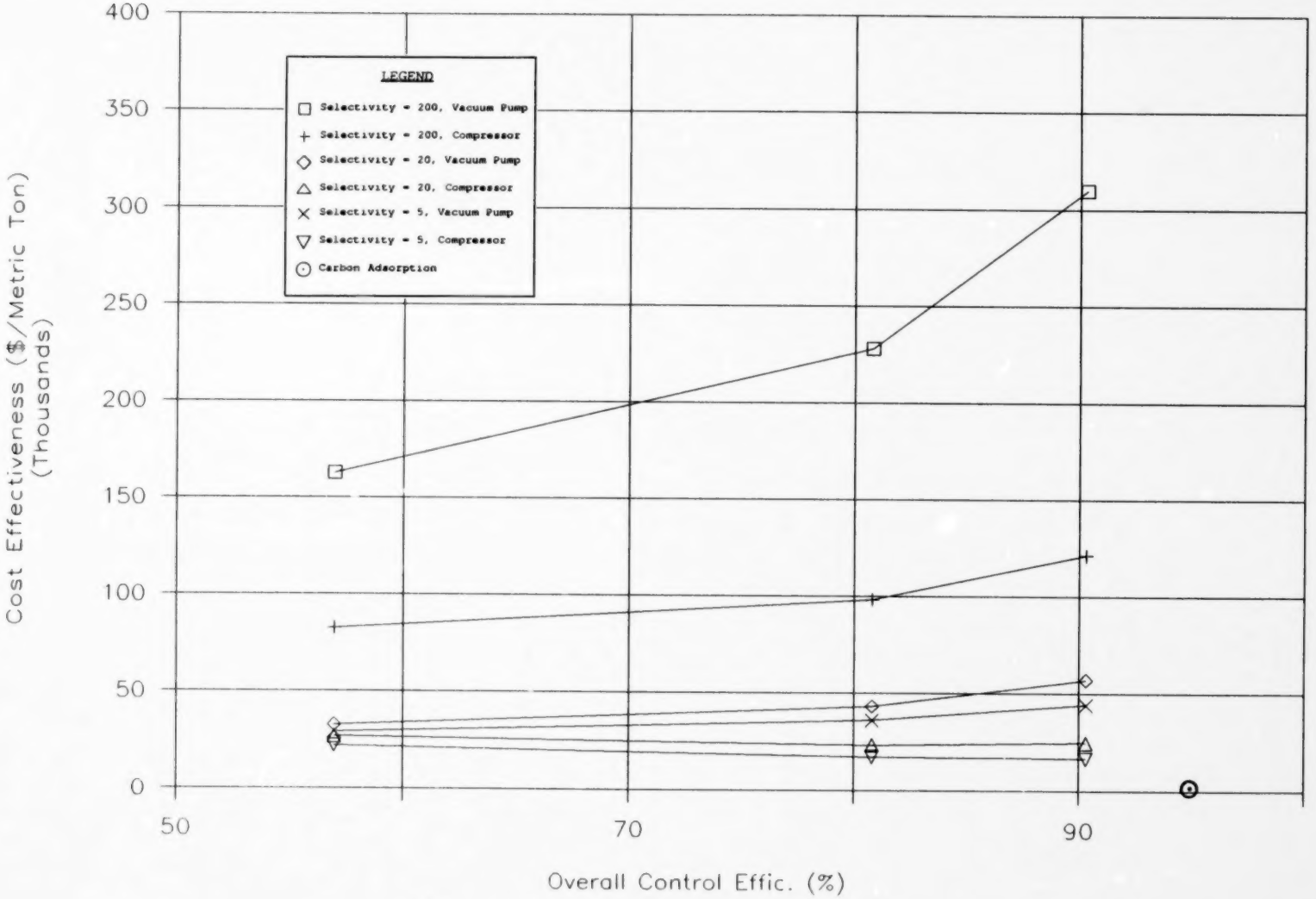

Figure 6-12. Annualized Cost Comparison (10000 ACFM, 100 ppm Toluene feed)

TABLE 6-5. EXAMPLE ANNUALIZED COST COMPARISON

System Flow Rate, Nm^3/s (cfm) = 62.2 (2500 ACFM) Inlet Solvent Concentration = 1000 ppm CFC-113	A. Carbon Adsorber Only[b]	B. Membrane Preconcent. w/Vac Pump[c]	C. Membrane Preconcent. w/Compr[d]
Total Capital Investment (From Table 6-3)	$164,100	$668,000	$396,800
Operating Days per Year[a]: 347			
Operating Hours per Day: 24			
Average Annual Operating Hours: 8328			
Direct Costs[b]			
Operating Labor	$9,800	$9,800	$9,800
Maintenance Labor	$3,800	$3,700	$3,700
Electricity	$1,100	$63,000	$117,300
Steam	$6,400	$3,200	$3,400
Cooling Water	$3,100	$1,600	$1,700
Wastewater Disposal	$9,400	$4,700	$5,000
Carbon Replacement	$1,000	$300	$300
Membrane Replacement	Not Applic.	$86,600	$30,000
TOTAL DIRECT COSTS	$34,600	$172,900	$171,200
Indirect Costs[b]			
Capital Recovery	$26,700	$108,700	$64,600
TOTAL INDIRECT COSTS	$26,700	$108,700	$64,600
Credits[b]			
Credit for Recovered Solvent	($33,500)	($19,900)	($19,800)
TOTAL ANNUALIZED OPERATING AND MAINTENANCE EXPENSES	$27,800	$261,700	$216,000
ANNUALIZED COST PER METRIC TON (MT) OF SOLVENT CONTROLLED	$114	$1800	$1500

Notes:
[a] Assumes solvent recovery unit has 95 percent availability.
[b] See Table 6-4 for Unit Costs.

SECTION 7

RECOMMENDATIONS FOR FUTURE RESEARCH

Application of membrane technology to separation of solvent vapors from air streams has not been fully developed on the commercial scale. Other gas-phase membrane processes have been practiced commercially for many more years. Examples of these other processes include hydrogen recovery from refinery process streams, natural gas processing (CO_2 removal), and air separation. A few applications of membranes to organic vapor recovery have been marketed. One example is the recovery of gasoline vapors at gasoline bulk storage terminals. However, further penetration of membrane technology into recovery of volatile organic solvents appears to require more testing and improvements. Specifically, the improvements would be toward higher removal efficiencies at lower cost.

Testing of Bench-scale Membrane with Carbon Adsorber: Future research might include studies which combine operational testing of a membrane device with a carbon adsorber. Although the economic analysis in this work showed that the membrane approach was uniformly more expensive, further work should be performed in the low concentration range (i.e., 20 - 100 ppmv). In this range, carbon adsorbers require larger beds and more frequent regeneration because the working capacity is lower. That is, the driving force for solvent adsorption becomes very low with dilute inlet concentrations.

Working capacities are usually estimated by carbon adsorber vendors based on past experience. The most accurate estimates are obtained by testing full-scale systems adsorbing the same compound(s). The next-best estimate is testing of a bench-scale adsorber operating at the same conditions as a proposed full-scale system. This report relied on a simplified approach used by carbon adsorber vendors which assumes that working capacity is generally half of the equilibrium capacity. While this simplified approach takes the

inlet concentration into account (via the equilibrium capacity), it may overstate the working capacity when dealing with low inlet concentration.

Therefore, a suggested research effort would be to combine a bench-scale membrane device with a bench-scale carbon adsorber for clean-up of dilute solvent streams. This test would be intended to determine if the membrane provides sufficient enrichment to allow improved working capacity of the carbon adsorber, resulting in reduced bed area and reduced steam regeneration demand.

Additional bench-scale testing should be performed with solvent vapors in air (instead of dry nitrogen only) to determine if the presence of oxygen and water have any effect on membrane performance. These results would be important for flammable solvents, since oxygen enrichment in the permeate could increase the risk of formation of explosive mixtures. The effect of water vapor is important, too, since humidity in the ambient air will result in water vapor as a normal constituent in membrane feed gases. The selectivity of a membrane towards water vapor will determine whether the permeate product is dry or wet.

Improved Membrane Materials: Another area for future research should be aimed at developing improved membrane materials, especially the active layer. With the current membrane, a compromise is struck between using a thin membrane allowing improved flux rates and solvent removal, but at the cost of poor selectivity and enrichment. This results is lower permeate concentrations and increased gas flow through the vacuum pump and carbon adsorber. Alternatively, one can specify a highly selective membrane, which will improve the enrichment. However, this thicker membrane will require a larger area to achieve the same degree of solvent removal.

Thus, a suggestion for future research would be to focus on testing of new membrane materials which exhibit both improved solvent permeability (to increase the solvent flux rate) and improved selectivity (to increase the enrichment). It is felt that most membrane vendors (e.g., MTR, Grace, Nitto Denko) are actively working in this area, but their research is proprie-

tary. University programs in polymer science also are studying membrane materials.

Alternative Arrangements: A third area for future research would be developing improved alternative arrangements for membrane devices. The tests conducted for this work used a simple one-pass arrangement. Other configurations include pressurizing the feed gas (which was economically analyzed, but not tested), and routing a small amount of residue gas to backflush the permeate side. Both of these options have potential advantages which should be tested further.

Furthermore, the hollow fiber design has been touted as superior to spiral wound modules in terms of packing density and selectivity (Nitto Denko has indicated they would be willing to supply us with a hollow fiber test module). Also, the "continuous column" design studied by University of Cincinnati merits further review. The continuous column is particularly interesting because the design concept tries to optimize removal efficiency with pressure ratio. Computer simulations indicate that high removals (95 percent or greater) can be obtained with relative ease with the continuous column design.

Industrial Application

The application of a membrane system to an industrial VOC emission source will be highly dependent on the situation. The cost of the system will probably be more expensive than conventional VOC controls, such as carbon adsorption or incineration, at least for the situations examined in this report. If newer, more selective, and high flux rate materials can be developed, the costs for systems may become more competitive.

Industrial applications that currently seem best suited for this technology are those that require a high quality recovered product or possibly a situation where activated carbon may not apply. For example, the recovery of ketones and aldehydes with activated carbon has resulted in bed fires that could potentially destroy the recovery system. With the polymeric membrane,

that problem will not exist; that is, unless activated carbon is used to recover the permeate vapors. A similar situation exists for 1,1,1-trichloroethane where activated carbon systems can decompose the molecule resulting in adverse by-products, including hydrochloric acid. Also, compounds, such as styrene, that become reactive during the high temperature steam regeneration of an activated carbon system will be recoverable with a membrane.

In summary, at the current time, the use of membrane systems for recovery at low concentrations will be expensive in comparison to activated carbon or even incineration. Improvements in the future could change this situation. In particular, improvements in better polymeric membranes at low costs and use of compressor systems over vacuum ring pumps might be required.

SECTION 8

References

1. Weller, S., and W.A. Steiner. Franctional Permeation Through Membranes. Chem. Eng. Prog., (46):585-591, 1950.

2. Pan, C.Y. and H.W. Habgood. An Analysis of the Single Stage Gaseous Permeation Process. Ind. Eng. Chem. Fundam., (13):323-331, 1974.

3. Peinemann, K.V., J.M. Mohr, and R.W. Baker. Separation of Organic Vapors from Air. AIChE Symp. Drt. 250(82):19, 1986.

4. Baker, R.W., N. Yoshioka, J.M. Mohr, and A.J. Khan. Separation of Organic Vapors from Air. J. Memb. Sci. (31):259-271, 1987.

5. Baker, R.W., I. Blume, V. Helm, A Khan, J. Maquire, and N. Yoshioka. Membrane Reearch in Energy and Solvent Recovery from Industrial Effluent Streams. DE 84/01/6819, Progress Report for the period November 5, 1982 - November 5, 1983, prepared by Membrane Technology & Research, Inc. for U.S. Department of Energy (INEL), 1984. 57 pp.

6. Armstrong, D., R.W. Bakr, and J. Mohr. New Membrane Preconcentrator Device for Trace Vapor Detection Systems; Phase I. AD-A161 042/7/XAB, prepared by Membrane Technology & Research, Inc. for U.S. Army Belvoir Research and Development Center, 1985. 23 p.p.

7. U.S. Patent No. 4,553, 983 to R.W. Baker and Membrane Technology and Research, Inc. Process for Recovering Organic Vapors from Air. Granted Nov. 19, 1985.

8. Strathman, H., C.M. Bell, and K. Kimmerle. Development of Synthetic Membranes for Gas and Vapor Separation. Pure & Appl. Chem. 58(12):1663-1668, 1986.

9. Ryans, J.L., and S. Croll. "Selecting Vacuum Systems." Chemical Engineering, 88(25):72-90, Dec. 14, 1981.

10. Capital and Operating Costs of Selected Air Pollution Control Systems - I. R.B. Neveril et al. Journal of the Air Pollution Control Association. Vol. 28, No. 8. August 1978.

11. The Cost of Controlling Organic Emissions. Kittleman and Akell. Chemical Engineering Progress, 74(4):87-91, April 1978.

12. Plant Design and Economics for Chemical Engineers (3rd Ed.). Peters and Timmerhaus. 1980.

13. Spillman, R.W. Economics of Gas Separation Membranes. Chem. Eng. Prog., (85):41-62, 1989.

14. Chen, S., Y.K. Kao, and S.T. Hwang. A Continuous Membrane Column Model Incorporating Axial Diffusion Terms. J. Memb. Sci., 26(2):143-164, 1986.

15. Hwang, S.T. and S. Ghalchi. Methane Separation by a Continuous Membrane Column. J. Memb. Sci. (11):187-198, 1982.

16. Hoover, K.C., and S.T. Hwang. Pervaporation by a Continuous Membrane Column. J. Memb. Sci. (10):253-271, 1982.

APPENDICES

APPENDIX A
TEST RESULTS

PARADOX RELATIONAL DATABASE

To simplify the data reduction of the raw experimental data, a relational database was chosen to manipulate the data. Table A-1 presents a listing of the laboratory notebook data which was entered into PARADOX.

Sample Calculations

Examples of the calculations performed to convert the raw data into usable results is shown in Figure A-1.

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TABLE A-1. ARCHIVED LOGBOOK DATA

Solvent	Date	Time	InFlo	InPr	InTmp	I	A	InAttn	OutPres	OutFlo	Vac	O.Area	OutAttn	SatTmp	SatFlo	SatPres	N2 Sup	Prm Ar	Prm Attn	1.Rf	O.Rf	PrmRf
CFC-113	1/09/89	13:52	55.0	21.7	71.0	8		50	1.5	47.5	26.8			68.0			76			1.00	1.00	1.00
CFC-113	1/09/89	14:00	55.0	21.8	72.0				2.0	47.5	26.8	9.0	50	68.0			76			1.00	1.00	1.00
CFC-113	1/09/89	14:09	55.0	21.4	72.1	8		50	2.0	47.5	26.8			68.0	5	60	69			1.00	1.00	1.00
CFC-113	1/09/89	14:26	55.0	21.0	72.7	47		50	1.8	47.5	26.8			68.0	47	35	73			1.00	1.00	1.00
CFC-113	1/09/89	14:34	57.0	22.4	73.2	76		50	3.0	49.0	26.8			74.0	73	34	68			1.00	1.00	1.00
CFC-113	1/09/89	14:45	57.0	21.6	73.0	65		50	1.2	49.0	26.8			80.0	73	34	73			1.00	1.00	1.00
CFC-113	1/09/89	14:53	57.0	21.9	73.1	69		50	1.4	49.0	26.8			83.0	73	34	74			1.00	1.00	1.00
CFC-113	1/09/89	15:05	56.0	21.5	73.6	65		50	1.5	48.0	26.8			86.0	73	34	60			1.00	1.00	1.00
CFC-113	1/09/89	15:20	55.5	21.2	74.1				.9	48.0	26.8	42.0	50	89.0	70	34	65			1.00	1.00	1.00
CFC-113	1/09/89	15:25	55.5	21.4	74.2				1.6	47.5	26.8	48.0	50	90.0	70	34	68			1.00	1.00	1.00
CFC-113	1/09/89	15:30	55.5	21.3	74.1				2.0	47.5	26.8	56.0	50	90.0	68	34	72			1.00	1.00	1.00
CFC-113	1/09/89	15:35	55.5	21.3	73.5				2.2	47.5	26.8	39.0	50	90.0	68	34	72			1.00	1.00	1.00
CFC-113	1/09/89	15:40	55.5	21.4	73.3				1.6	48.0	26.8	38.0	50	90.0	68	34	72			1.00	1.00	1.00
CFC-113	1/09/89	15:45	55.5	21.3	73.6	70		50	1.2	47.5	26.8			90.0	68	34				1.00	1.00	1.00
CFC-113	1/09/89	15:50	55.5	21.6	73.9	74		50	1.5	48.0	26.8			91.0	70	34	65			1.00	1.00	1.00
CFC-113	1/09/89	15:55	55.5	21.3	73.4	71		50	2.0	48.0	26.8			90.0	[illegible]	34	68			1.00	1.00	1.00
CFC-113	1/09/89	16:00	56.0	21.2	73.6	62		50	1.7	47.5	26.8			90.0	69	34	72			1.00	1.00	1.00
CFC-113	1/09/89	16:05	56.0	21.1	73.6	60		50	1.7	47.5	26.8			90.0	69	34	72			1.00	1.00	1.00
CFC-113	1/09/89	16:15	55.5	21.0	74.0				2.0	48.0	26.8			91.0	68	34	72			1.00	1.00	1.00
CFC-113	1/09/89	16:20	55.5	21.4	73.4				1.5	48.0	26.8			90.0	68	34	68			1.00	1.00	1.00
CFC-113	1/09/89	17:00	55.5	21.3	71.0	56		100	1.0	48.0	26.8			97.0	100	40	65			1.04	1.00	
CFC-113	1/09/89	17:10	56.0	21.1	72.0	24		500	1.5	48.0	26.8			104.0	100	42	69			1.04	1.00	
CFC-113	1/09/89	17:20	56.0	21.2	72.8	43		500	1.7	48.0	26.8			108.0	100	38	58			1.04	1.00	
CFC-113	1/09/89	17:25	56.0	21.2	73.0	42		500	2.3	48.0	26.8			110.0	100	37	70			1.04	1.00	
CFC-113	1/09/89	17:40	56.0	21.0	73.3	41		500	.8	48.0	26.8			113.0	100	37	60			1.04	1.00	
CFC-113	1/09/89	17:45	56.0	21.0	73.5	43		500	1.2	48.0	26.8			114.0	100	37	65			1.04	1.00	
CFC-113	1/09/89	17:55	56.0	20.0	73.8				.5	48.0	26.8	61.0	200	115.0	100	37	70			1.04	1.00	
CFC-113	1/09/89	19:20	55.0	21.0	72.4	60		500	1.5	47.5	27.0			119.0	17	44	75			1.04	1.00	
CFC-113	1/09/89	19:25	56.0	22.0	72.1	60		500	2.5	49.0	26.9			120.0	20	46	65			1.04	1.00	
CFC-113	1/09/89	19:30	55.5	21.9	71.7	58		500	2.3	47.5	26.9			120.0	20	47	64			1.04	1.00	
CFC-113	1/09/89	19:35	56.0	21.0	71.9				1.3	48.0	26.9	81.0	200	119.0	21	48	72			1.04	1.00	
CFC-113	1/09/89	19:40	56.0	21.6	72.2				2.0	48.0	26.9	83.0	200	119.0	21	49	63			1.04	1.00	
CFC-113	1/09/89	19:45	56.0	21.3	72.3	62		500	1.8	48.0	26.8			119.0	22	49	71			1.04	1.00	
CFC-113	1/09/89	19:50	56.0	21.2	72.6	63		500	1.0	48.0	26.8			119.0	22	50	71			1.04	1.00	
CFC-113	1/09/89	19:58	56.0	21.7	73.0				2.0	48.0	26.8	86.0	200	119.0	22	50	65			1.04	1.00	
CFC-113	1/09/89	20:03	56.0	21.2	73.0				1.8	48.0	26.8	86.0	200	119.0	22	50	75			1.04	1.00	
CFC-113	1/09/89	20:06	56.0	21.7	73.3	63		500	1.9	48.0	26.8			119.0	22	50	57					
CFC-113	1/09/89	20:10	56.0	21.9	73.6	64		500	2.0	48.0	26.8			119.0	25	50						
CFC-113	1/10/89	10:52	55.0	22.0	69.6	58		5000	2.9	48.0	26.9			146.0	92	34	65			1.04	1.09	
CFC-113	1/10/89	10:55	55.5	21.5	70.0	58		5000	1.0	48.0	26.8			146.0	95	34	68			1.04	1.09	
CFC-113	1/10/89	11:00	55.5	21.7	70.7				1.8	48.0	26.8	73.0	2000	147.0	97	34	69			1.04	1.09	
CFC-113	1/10/89	11:07	56.0	21.3	71.2				1.1	48.5	26.8	74.0	2000	147.0	93	34	73			1.04	1.09	
CFC-113	1/10/89	11:10	56.0	21.3	71.6	57		5000	.5	48.0	26.8			147.0	95	34	61			1.04	1.09	
CFC-113	1/10/89	11:13	56.0	21.4	71.6	59		5000	2.1	48.5	26.8			147.0	93	34	61			1.04	1.09	
CFC-113	1/10/89	11:17	56.0	21.3	71.3	60		5000	1.8	48.5	26.8			147.0	93	34	65			1.04	1.09	
CFC-113	1/10/89	11:22	56.0	21.2	71.6				.5	48.5	26.8	80.0	2000	148.0	93	34	68			1.04	1.09	
CFC-113	1/10/89	11:26	56.0	21.5	71.7				1.8	48.5	26.8	80.5	2000	148.0	93	34	62					
CFC-113	1/10/89	11:29	56.0	21.5	72.0				1.5	49.0	26.8	80.0	2000	147.0	93	34	62			1.04	1.09	
CFC-113	1/10/89	11:35	56.0	21.4	72.4	58		5000	1.5	48.5	26.8			147.0	95	34	65			1.04	1.09	
CFC-113	1/10/89	11:40	56.0	21.0	72.6	57		5000	2.0	48.0	26.8			148.0	93	35	60			1.04	1.09	

TABLE A-1. (CONTINUED)

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Solvent	Date	Time	InFlo	InPr	InTmp	I	A	InAttn	OutPres	OutFlo	Vac	O.Area	OutAttn	SatTmp	SatFlo	SatPres	N2 Sup	Prm Ar	Prm Attn	I.Rf	O.Rf	PrmRf
CFC-113	1/10/89	11:45	56.0	21.2	72.7	57		5000	1.2	48.0	26.8			148.0	90	34	70			1.04	1.09	
CFC-113	1/10/89	11:50	56.0	21.2	72.8				1.1	48.5	25.8	76.0	2000	148.0	90					1.04	1.09	
CFC-113	1/10/89	11:55	56.0	21.7	73.1				2.3	48.0	26.8	76.0	2000	148.0	90	34	64			1.04	1.09	
CFC-113	1/10/89	11:58	56.0	21.5	73.3				2.0	48.5	26.9	72.0	2000	148.0	90	34	72			1.04	1.09	
CFC-113	1/10/89	12:00	56.0	21.2	73.4				1.4	48.0	26.8	69.0	2000	148.0	90	34	71			1.04	1.09	
CFC-113	1/10/89	12:06	56.0	21.8	73.4	55		5000	1.6	48.5	26.8			148.0	90	34	71			1.04	1.09	
CFC-113	1/10/89	12:11	56.0	21.5	72.8	54		5000	1.1	48.5	26.8			148.0	90	35	71			1.04	1.09	
CFC-113	1/10/89	12:16	56.0	21.2	72.9				9.0	48.5	26.8	70.0	2000	148.0	90	35	72			1.04	1.09	
CFC-113	1/10/89	12:20	56.0	21.5	73.1				1.0	48.5	26.7	71.0	2000	148.0	90	35	64			1.04	1.09	
CFC-113	1/10/89	12:24	56.0	21.2	73.2	56		5000	1.0	48.0	26.8			148.0	90	35	68			1.04	1.09	
Hexane	12/14/88	16:15	55.0	19.8	72.3	52		5000	.5	46.5	27.7			73.0	60	35	37			1.37	1.37	1.37
Hexane	12/14/88	16:19	55.0	20.0	72.8				1.0	47.5	27.7	51.0	2000	73.0	65	35	37			1.37	1.37	1.37
Hexane	12/14/88	16:29	55.0	20.0	71.9				1.0	47.0	27.7	49.0	2000	73.0	63	35	37			1.37	1.37	1.37
Hexane	12/14/88	16:39	55.0	20.0	72.2	57		5000	1.0	47.0	27.7			72.0	63	35	37			1.37	1.37	1.37
Hexane	12/14/88	16:45	57.0	22.0	73.0	53		5000	1.0	49.0	27.7			72.0	60	35	37			1.37	1.37	1.37
Hexane	12/14/88	16:50	56.0	21.6	72.4	53		5000	1.5	48.0	27.7			72.0	60	35	37			1.37	1.37	1.37
Hexane	12/14/88	17:00	56.0	21.6	72.2				1.5	48.0	27.7	49.5	2000	72.0	60	35	37			1.37	1.37	1.37
Hexane	12/14/88	17:05	55.5	20.5	72.8				0.0	48.0	27.7	48.0	2000	72.0	60	35	37			1.37	1.37	1.37
Hexane	12/14/88	17:15	55.5	21.2	72.8	54		5000	1.7	48.0	27.7			72.0	60	35	37			1.37	1.37	1.37
Hexane	12/14/88	17:55	60.0	23.2	72.9	45		50	1.5	52.0	27.6			101.0		35	37			1.23	1.38	
Hexane	12/14/88	18:15	58.5	22.9	73.2	32		50	.5	51.0	27.7			111.0		35	37			1.23	1.38	
Hexane	12/14/88	18:45	54.0	20.0	72.8	29		20000	2.0	46.0	27.7			128.0	100	35	37			1.23	1.38	
Hexane	12/14/88	19:15	56.0	21.0	72.4	47		20000	1.0	47.5	27.7			146.0	100	35	37			1.23	1.38	
Hexane	12/14/88	19:35	55.5	20.2	73.5	59		20000	1.0	48.0	27.7			149.0	100	35	37			1.23	1.38	
Hexane	12/14/88	19:45	57.0	21.5	72.6	53		20000	1.0	49.0	27.6			149.0		35	37			1.23	1.38	
Hexane	12/14/88	20:00	56.5	21.1	73.8				.5	48.5	27.5	44.0	10000	150.0		35	37			1.23	1.38	
Hexane	12/14/88	20:10	56.0	21.1	73.2	57		20000	1.5	47.5	27.4			150.0		35	37			1.23	1.38	
Hexane	12/14/88	20:15	56.0	21.1	72.8	57		20000	1.7	47.5	27.4			150.0		35	37			1.23	1.38	
Hexane	12/14/88	20:25	56.0	21.1	73.5				1.5	48.0	27.3	48.0	10000	149.0		35	37			1.23	1.38	
Hexane	12/14/88	20:30	56.0	21.1	73.8				1.2	48.0	27.2	48.0	10000	149.0		35	37			1.23	1.38	
Hexane	1/20/89	16:15	56.0	21.8	72.4	31		50	1.8	49.0	26.8			69.0			43			1.34	1.34	1.34
Hexane	1/20/89	16:19	56.0	22.0	72.4	28		50	.9	48.5	26.8			69.0			40			1.34	1.34	1.34
Hexane	1/20/89	16:27	56.0	21.2	72.5				0.0	48.5	26.8	26.0	50	69.0			46			1.34	1.34	1.34
Hexane	1/20/89	16:31	56.0	21.8	72.5				1.0	48.5	26.8	25.0	50	69.0			44			1.34	1.34	1.34
Hexane	1/20/89	16:50	56.0	21.7	72.4	94		500	1.0	48.5	26.8			69.0	12	41	39			1.34	1.34	1.34
Hexane	1/20/89	17:00	56.0	21.3	71.7	50		1000	.9	48.5	26.8			69.0	13	42	41	76	2000	1.34	1.34	1.34
Hexane	1/20/89	17:05	56.0	21.5	71.8	50		1000	1.2	48.5	26.8			69.0	13	42	43	76	2000	1.34	1.34	1.34
Hexane	1/20/89	17:15	56.0	21.2	72.0				1.8	48.0	26.8	50.0	500	69.0	13	41	45	76	2000	1.34	1.34	1.34
Hexane	1/20/89	17:20	56..	21.3	72.0				1.8	48.0	26.8	47.5	500	69.0	13	41	43	76	2000	1.34	1.34	1.34
Hexane	1/20/89	17:25	56.0	21.4	72.1				1.2	48.0	26.8	49.5	500	69.0	13	41	42	76	2000	1.34	1.34	1.34
Hexane	1/20/89	17:32	56.0	21.7	71.5	52		1000	1.1	48.0	26.8			69.0	13	41	42	76	2000	1.34	1.34	1.34
Hexane	1/20/89	17:37	56.0	21.1	71.5	52		1000	.7	48.0	26.8			69.0	13	41	46	76	2000	1.34	1.34	1.34
Hexane	1/20/89	17:42	56.0	21.1	71.7	52		1000	1.6	48.0	26.8			69.0	13	41	46	76	2000	1.34	1.34	1.34
Hexane	1/20/89	17:50	56.0	21.0	72.0				.6	48.5	26.8			69.0	13	41	43	76	2000	1.34	1.34	1.34
Hexane	1/20/89	17:55	56.0	21.9	71.5				2.5	48.5	26.8			69.0	13	41	38	76	2000	1.34	1.34	1.34
Hexane	1/20/89	18:04	56.0	21.7	71.5	53		1000	1.5	48.5	26.8			68.0	12	41	39	76	2000	1.34	1.34	1.34
Hexane	1/20/89	18:10	56.0	21.3	71.6	53		1000	1.9	48.5	26.8		500	68.0	12	41	42	76	2000	1.34	1.34	1.34
Hexane	1/20/89	18:18	56.0	21.5	72.0				1.0	48.5	26.8	52.5	500	68.0	12	41	42	76	2000	1.34	1.34	1.34
Hexane	1/20/89	18:23	56.0	21.3	71.6				1.2	48.5	26.8	51.5	500	69.0	12	41	43	76	2000	1.34	1.34	1.34

TABLE A-1. (CONTINUED)

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Solvent	Date	Time	InFlo	InPr	InTmp	I	A	InAttn	OutPres	OutFlo	Vac	O.Area	OutAttn	SatTmp	SatFlo	SatPres	N2 Sup	Prm Ar	Prm Attn	I.Rf	O.Rf	PrmRf
Hexane	1/20/89	18:32	56.0	21.1	71.3	53		1000	1.0	48.5	26.8			73.0	12	42	46					
Hexane	1/20/89	18:40	56.0	21.2	71.3	38		10000		48.5	26.8			86.0	72	38	41					
Hexane	1/20/89	18:47	56.0	21.5	71.1	56		10000	.6	49.0	26.8			96.0	92	35	36					
Hexane	1/20/89	18:53	56.0	21.5	71.2	56		10000	1.5	49.0	26.8			98.0	90	35	38	79	20000	1.33	1.33	1.19
Hexane	1/20/89	19:00	56.0	21.5	71.5				.6	49.0	26.8	56.0	5000	100.0	90	35		79	20000	1.33	1.33	1.19
Hexane	1/20/89	19:05	56.0	21.8	71.6				1.5	49.0	26.8	58.0	5000	100.0	90	35	40	79	20000	1.33	1.33	1.19
Hexane	1/20/89	19:10	56.0	22.0	71.2				1.6	49.0	26.8	57.0	5000	101.0	90	35	40					
Hexane	1/20/89	19:15	56.0	21.5	71.0	57		10000	1.6	49.0	26.8			101.0	90	35	42	79	20000	1.33	1.33	1.19
Hexane	1/20/89	19:20	56.5	21.3	71.1	57		10000	.7	49.0	26.8			100.0	90	35		79	20000	1.33	1.33	1.19
Hexane	1/20/89	19:25	56.5	21.3	71.3	58		10000	.4	49.0	26.8			100.0	90			79	20000	1.33	1.33	1.19
Hexane	1/20/89	19:34	56.0	21.2	71.3				.3	49.0	26.8			101.0	90	35	36	79	20000	1.33	1.33	1.19
Hexane	1/20/89	19:42	56.5	21.3	71.0				.5	49.0	26.8	60.0	5000	100.0	90	35	36	79	20000	1.33	1.33	1.19
Hexane	1/20/89	19:44	56.5	21.5	71.2				1.2	49.0	26.8	59.0	5000	101.0	90	35	37	79	20000	1.33	1.33	1.19
Hexane	1/20/89	19:50	56.5	21.5	71.3				1.0	49.0	26.8	57.5	5000	101.0	90	35	37	79	20000	1.33	1.33	1.19
MEK	12/16/88	14:20	56.0	21.0	68.8	73		50	1.2	48.0	28.1			131.0		35	37			1.23	1.38	
MEK	12/16/88	14:40	55.5	2.0	70.6	81		50	1.0	48.0	28.1	80.0	50	140.0		35	37			1.23	1.38	
MEK	12/16/88	14:55	56.0	21.0	72.0				1.0	48.0	28.1	72.0	50	146.0		35	37			1.23	1.38	
MEK	12/16/88	15:40	60.0	24.0	73.1	39		50	1.0	53.0	28.1			150.0		35	37			1.23	1.38	
MEK	12/16/88	15:50	58.0	21.2	73.4	49		1000	3.0	49.5	28.1			153.0	25	35	37			1.23	1.38	
MEK	12/16/88	16:00	56.5	21.2	74.1	49		1000	2.0	48.5	28.1			148.0	27	35	37			1.23	1.38	
MEK	12/16/88	16:06	57.0	21.8	73.2	22		1000	1.8	49.0	28.1			146.0	35	35	37			1.23	1.38	
MEK	12/16/88	16:18	58.0	21.9	73.8	51		500	1.0	50.0	28.1			148.0	37	35	37			1.23	1.38	
MEK	12/16/88	16:23	57.0	21.9	74.1				1.0	49.5	28.1	53.0	200	150.0	37	35	37			1.23	1.38	
MEK	12/16/88	16:35	56.0	21.6	74.4				1.5	48.5	28.1	80.0	100	153.0	37	35	37			1.23	1.38	
MEK	12/16/88	16:45	57.0	21.9	73.5	54		20000	1.0	49.5	28.1			153.0	37	35	37			1.23	1.38	
MEK	12/16/88	16:53	57.5	21.8	74.1	59		20000	.5	49.5	28.1			153.0	37	35	37			1.23	1.38	
MEK	12/16/88	17:07	56.0	21.7	74.3				1.0	48.5	28.1	53.0	200	152.0	37	35	37			1.23	1.38	
MEK	12/16/88	17:12	56.0	21.7	73.8				1.3	48.5	28.1	51.5	200	151.0	37	35	37			1.23	1.38	
MEK	12/16/88	17:27	56.0	21.1	74.2	64		20000	1.0	48.0	28.0			150.0	37	35	37			1.23	1.38	
MEK	12/16/88	17:36	56.0	21.0	74.7	67		20000	.5	48.0	28.0			150.0	37	35	37			1.23	1.38	
MEK	12/16/88	17:57	56.0	21.1	73.4				1.5	48.0	28.1	60.0	200	149.0	37	35	37			1.23	1.38	
MEK	12/16/88	18:05	56.0	21.1	73.8				1.0	48.0	28.1	53.0	200	149.0	37	35	37			1.23	1.38	
MEK	12/16/88	18:10	56.0	21.0												35	37			1.23	1.38	
MEK	12/16/88	18:16	56.5	21.6	74.4	87		20000	1.3	48.5	28.1					35	37			1.23	1.38	
MEK	12/16/88	18:30	56.5	21.3	74.8	**		20000	.8	48.5	28.1					35	37			1.23	1.38	
MEK	12/19/88	18:47	56.0	21.5	73.1	15		50	2.0	49.0	26.8			175.0		30	37			1.23	1.38	
MEK	12/19/88	19:02	55.0	20.9	72.4				2.5	47.5	26.8	12.0	50	178.0	0	33	37			1.08	1.00	
MEK	12/19/88	19:13	55.5	21.7	73.8	23		2000	2.0	48.0	26.8			181.0	35	32	37			1.08	1.00	
MEK	12/19/88	19:18	56.0	21.6	74.0	26		2000	1.2	48.5	26.8			184.0	55	32	37			1.08	1.00	
MEK	12/19/88	19:27	55.5	21.4	73.4	51		2000	1.0	48.0	26.8			185.0	85	32	37			1.08	1.00	
MEK	12/19/88	19:32	56.0	21.2	74.0	69		2000	2.0	48.5	26.8			185.0	85	32	37			1.08	1.00	
MEK	12/19/88	19:45	55.5	21.1	74.4	83		2000	2.0	48.0	26.8			184.0	80	32	37			1.08	1.00	
MEK	12/19/88	19:50	56.0	21.5	74.0	79		2000	1.5	49.0	26.8			183.0	80	32	37			1.08	1.00	
MEK	12/19/88	19:55	56.5	21.6	73.6	78		2000	1.0	49.0	26.8			183.0	80	2	37			1.08	1.00	
MEK	12/19/88	20:05	56.5	21.4	74.5				1.5	49.0	26.8	64.0	1000	182.0	80	32	37			1.08	1.00	
MEK	12/19/88	20:10	56.5	21.4	75.0				.1	49.0	26.8	63.0	1000	182.0	80	32	37			1.08	1.00	
MEK	12/19/88	20:16	55.5	21.5	74.2	73		2000	1.0	48.5	26.8			182.0	80	32	37			1.08	1.00	
MEK	12/19/88	20:27	55.0	21.0	74.0	75		2000	1.5	47.5	26.8			181.0	80	32	37			1.08	1.00	
MEK	12/19/88	20:40	56.0	21.5	75.0				1.0	48.5	26.8	67.0	1000	180.0	80	32	37			1.08	1.00	

TABLE A-1. (CONTINUED)

Solvent	Date	Time	InFlo	InPr	InTmp	I	A	InAttn	OutPres	OutFlo	Vac	O.Area	OutAttn	SatTmp	SatFlo	SatPres	N2 Sup	Prm Ar	Prm Attn	I.Rf	O.Rf	PrmRf
MEK	12/19/88	20:45	56.0	21.3	74.2				1.5	48.5	26.8	64.0	1000	181.0	80	32	37			1.08	1.00	
MEK	12/19/88	20:50	56.0	21.3	73.7				1.5	48.5	26.8	62.0	1000	180.0	80	32	37			1.08	1.00	
MEK	12/19/88	21:00	56.0	21.2	74.5	71		2000	2.0	48.0	26.8			180.0	80	32	37			1.08	1.00	
MEK	12/19/88	21:05	56.0	21.2	74.8	76		2000	1.0	48.5	26.8			180.0	80	32	37			1.08	1.00	
MEK	12/19/88	21:10	56.0	21.2	7.5	80		2000	1.5	49.0	26.8			180.0	80	32	37			1.08	1.00	
MEK	12/19/88	21:15	56.0	21.2	73.8				1.5	49.0	26.8	80.0	1000	180.0	80	32	37			1.08	1.00	
MEK	12/19/88	21:20	56.0	21.2	73.9				1.5	48.5	26.8	70.0	1000	180.0	80	32	37			1.08	1.00	
MEK	12/19/88	21:25	56.5	21.6	74.3				1.0	49.0	26.8	68.0	1000	180.0	80	32	37			1.08	1.00	
MEK	12/19/88	21:30	56.0	21.5	74.7				.5	48.5	26.8	67.0	1000	180.0	80	32				1.08	1.00	
MEK	12/19/88	21:40	56.5	21.3	74.0	74		2000	1.2	49.0	26.8			180.0	80	32	37			1.08	1.00	
MEK	12/19/88	21:45	56.5	21.6	73.5	73		2000	2.0	49.0	26.8			180.0	80	32	37			1.08	1.00	
MEK	12/19/88	21:60	56.0	21.5	74.7				.5	48.5	26.8	67.0	1000	180.0	80	32	37			1.07	1.18	
MEK	1/03/89	18:40	55.0	21.1	71.2	58		100	1.3	48.0	26.9			196.0		40	33			1.07	1.18	
MEK	1/03/89	19:00	56.0	21.7	73.1				1.3	48.0	26.9	92.0	50	196.0		40	31			1.07	1.18	
MEK	1/03/89	19:05	56.0	21.4	73.0	77		50	.5	48.5	26.8			196.0		40	32			1.07	1.18	
MEK	1/03/89	19:30	57.0	22.1	74.3	47		10000	1.7	50.0	26.8			207.0	30	37	31			1.07	1.18	
MEK	1/03/89	19:35	56.0	21.3	74.6	68		10000	1.0	48.0	26.8			207.0	30	36	32			1.07	1.18	
MEK	1/03/89	19:45	56.5	21.5	74.5	84		10000	2.0	49.0	26.8			207.0		42	26			1.07	1.18	
MEK	1/03/89	20:00	56.0	21.4	74.7	44		10000	3.2	49.0	26.8			204.0	100	48	32			1.07	1.18	
MEK	1/03/89	20:30	58.0	22.9	75.3	41		10000		50.0	26.8			204.0	100	24	32			1.07	1.18	
MEK	1/03/89	20:55	56.0	21.3	75.2	62		10000	2.1	49.0	26.8			201.0	100	24	31			1.07	1.18	
MEK	1/03/89	21:00	57.0	21.5	75.4	61		10000	1.5	49.0	26.8			201.0	100	24	31			1.07	1.18	
MEK	1/03/89	21:05	57.0	21.5	74.8	58		10000	1.0	49.0	26.8			200.0	100	24	31			1.07	1.18	
MEK	1/03/89	21:25	56.0	21.0	75.7				.5	48.0	26.8	63.0	5000	199.0	100	24	31			1.07	1.18	
MEK	1/03/89	21:30	57.0	21.6	74.9				1.0	49.0	26.8	61.0	5000	199.0	100	24	32			1.07	1.18	
MEK	1/03/89	21:35	57.0	21.5	75.0				1.0	49.0	26.8	61.0	5000	198.0	100	24	31			1.07	1.18	
MEK	1/03/89	21:40	57.0	21.9	75.3	77		10000		50.0	26.8			198.0	100	24	32			1.07	1.18	
MEK	1/03/89	21:45	56.0	21.0	75.6	77		10000	.5	49.0	26.8			198.0	100	24	32			1.07	1.18	
MEK	1/03/89	21:50	57.0	21.6	75.6	81		10000	1.6	49.0	26.8			198.0	100	24	33			1.07	1.18	
MEK	1/03/89	21:55	57.0	21.7	75.0	81		10000	1.0	49.0	26.8			198.0	100	24	32			1.07	1.18	
MEK	1/03/89	22:00	57.0	21.7	75.0	75		10000	1.0	49.0	26.8			198.0	100	24	32			1.07	1.18	
MEK	1/03/89	22:05	57.0	21.8	75.5				1.0	49.0	26.8	89.0	5000	198.0	100	24				1.07	1.18	
MEK	1/03/89	22:10	57.0	21.8	75.6				.5	49.0	26.8	89.0	5000	198.0	100	24	32			1.07	1.18	
MEK	1/03/89	22:15	56.0	21.5	75.3				1.0	49.0	26.8	80.0	5000	198.0	100	24	32			1.07	1.18	
MEK	1/03/89	22:20	56.0	21.5	74.9				1.5	49.0	26.8	88.0	5000	198.0	100	24	32			1.07	1.18	
MEK	1/03/89	22:25	56.0	21.5	75.1	80		10000												1.07	1.18	
MEK	1/03/89	22:30	56.0	20.9	75.3	83		10000		49.0	26.8			197.0	100	21	32			1.07	1.18	
MeCl	1/05/89	15:38	57.0	21.8	73.8				1.6	49.0	26.9	11.0	50	67.0			73			1.12	1.05	
MeCl	1/05/89	15:45	56.0	21.8	73.4				1.1	49.0	26.9	9.0	50	67.0			73			1.12	1.05	
MeCl	1/05/89	15:52	57.0	21.7	74.0	9		50	.5	50.0	26.9			67.0			76			1.12	1.05	
MeCl	1/05/89	16:05	56.0	21.6	74.1	75		200	2.5	49.0	26.9			67.0		28	75			1.12	1.05	
MeCl	1/05/89	16:30	55.0	21.8	74.2	55		200	1.3	48.0	26.8			68.0	5	64	50			1.12	1.05	
MeCl	1/05/89	16:35	56.0	21.4	75.0	55		200	1.3	48.0	26.8			67.0	5	63	59			1.12	1.05	
MeCl	1/05/89	17:01	55.0	21.6	75.1				3.0	48.0	26.8	51.0	100	68.0	5	68				1.12	1.05	
MeCl	1/05/89	17:06	56.0	21.6	75.0				2.4	48.0	26.8	51.0	100	68.0	5	68				1.12	1.05	
MeCl	1/05/89	17:25	56.0	21.4	75.2	57		200	1.5	48.0	26.8			68.0	5	69	67			1.12	1.05	
MeCl	1/05/89	17:30	56.0	21.7	75.2	56		200	1.7	48.0	26.8			68.0	5	69	75			1.12	1.05	
MeCl	1/05/89	17:40	55.0	21.5	75.4				2.0	47.0	26.8	51.0	100	68.0	5	69				1.12	1.05	
MeCl	1/05/89	17:45	56.0	21.5	75.2				2.1	48.0	26.8	50.0	100	68.0	5	69	68			1.12	1.05	

TABLE A -1. (CONTINUED)

Solvent	Date	Time	InFlo	InPr	InTmp	I	A	InAttn	OutPres	OutFlo	Vac	O.Area	OutAttn	SatTmp	SatFlo	SatPres	N2 Sup	Prm Ar	Prm Attn	I.Rf	O.Rf	PrmRf
MeCl	1/05/89	17:55	57.0	21.7	74.7	50		200	1.6	49.0	26.8			68.0	5	69	57			1.12	1.05	
MeCl	1/05/89	18:00	57.0	21.5	75.1	49		200	.5	49.0	26.8			68.0	5	69	63			1.12	1.05	
MeCl	1/05/89	18:40	56.0	21.7	75.5	48		200	2.6	48.0	26.8			68.0	5	68				1.12	1.05	
MeCl	1/05/89	18:48	56.0	21.5	75.1				2.0	48.0	26.8	42.0	100	68.0	5	68				1.12	1.05	
MeCl	1/05/89	18:51	56.0	21.5	75.2				1.5	48.0	26.8	42.0	100	68.0	5	68				1.12	1.05	
MeCl	1/05/89	18:58	56.0	21.5	75.6	47		200	1.5	48.0	26.8			68.0	5	68				1.12	1.05	
MeCl	1/05/89	19:03	56.0	21.8	75.1	46		200	2.0	48.0	26.8			68.0	5	68	75			1.12	1.05	
MeCl	1/05/89	19:14	55.0	21.3	75.0				1.0	48.0	26.8	42.0	100	68.0	5	69	74			1.12	1.05	
MeCl	1/05/89	19:18	56.0	21.5	75.5				2.0	48.0	26.8	41.0	100							1.12	1.05	
MeCl	1/06/89	13:50	55.0	21.5	72.2	24		50	1.9	48.0	26.8		100				70			1.10	1.09	1.08
MeCl	1/06/89	13:54	55.0	21.3	72.5	23		50	2.1	48.0	26.8		100	68.0			75			1.10	1.09	1.08
MeCl	1/06/89	14:00	55.0	21.1	73.5				2.5	48.0	26.8	22.0	50	67.0			75			1.10	1.09	1.08
MeCl	1/06/89	14:05	57.0	21.5	73.8				.5	49.0	26.8	21.0	50	68.0			67			1.10	1.09	1.08
MeCl	1/06/89	14:13	56.0	21.7	73.8	17		50	1.5	48.0	26.8			67.0			75			1.10	1.09	1.08
MeCl	1/06/89	14:17	56.0	21.5	74.2	16		50	1.5	48.0	26.8			67.0	5	37	66			1.10	1.09	1.08
MeCl	1/06/89	14:24	56.0	21.1	75.0	61		200	1.7	48.0	26.8			67.0	5	48	75			1.10	1.09	1.08
MeCl	1/06/89	14:27	56.0	21.7	75.2	84		100	1.5	48.0	26.8			68.0	5	50	63			1.10	1.09	1.08
MeCl	1/06/89	14:31	56.0	21.8	74.8	89		100	2.5	48.0	26.8			68.0	5	52	75	60	500	1.10	1.09	1.08
MeCl	1/06/89	14:35	56.0	21.7	74.6	48		200	2.0	48.0	26.8			68.0	5	53	75	60	500	1.10	1.09	1.08
MeCl	1/06/89	14:41	56.0	21.1	75.3	50		200	1.5	48.0	26.8			68.0	5	53	75	60	500	1.10	1.09	1.08
MeCl	1/06/89	14:45	56.0	21.4	75.6	52		200	2.1	48.0	26.8			68.0	5	54	75	60	500	1.10	1.09	1.08
MeCl	1/06/89	14:49	55.0	21.2	75.9	52		200	1.3	48.0	26.8			68.0	5	54	75	60	500	1.10	1.09	1.08
MeCl	1/06/89	14:54	55.0	21.3	75.5				2.6	48.0	26.8	43.0	100	68.0	5	54	75	60	500	1.10	1.09	1.08
MeCl	1/06/89	14:58	55.0	21.5	75.3				1.5	48.0	26.8	43.0	100	69.0	5	54	75	60	500	1.10	1.09	1.08
MeCl	1/06/89	15:05	55.0	21.2	75.6	53		200	2.1	48.0	26.8			68.0	5	54	74	60	500	1.10	1.09	1.08
MeCl	1/06/89	15:08	55.0	21.5	76.0	53		200	2.5	48.0	26.8			68.0	5	54	67	60	500	1.10	1.09	1.08
MeCl	1/06/89	15:15	55.0	21.5	76.1				2.6	48.0	26.8			68.0	5	54	70	60	500	1.10	1.09	1.08
MeCl	1/06/89	15:20	56.0	21.7	75.7				2.0	48.0	26.8			69.0	5	54	66	60	500	1.10	1.09	1.08
MeCl	1/06/89	15:25	56.0	21.6	75.6				2.1	48.0	26.8			68.0	5	54	69	60	500	1.10	1.09	1.08
MeCl	1/06/89	15:30	56.0	21.6	76.2				2.5	48.0	26.8			68.0	5	54		60	500	1.10	1.09	1.08
MeCl	1/06/89	15:38	56.0	21.5	76.2	51		200		48.0	26.8							60	500	1.10	1.09	1.08
MeCl	1/06/89	15:42	56.0	21.5	75.9	50		200	2.7	48.0	26.8			69.0	5	54	71	60	500	1.10	1.09	1.08
MeCl	1/06/89	15:47	55.0	21.6	75.4				2.6	47.0	26.8	43.0	100	69.0	5	54	70	60	500	1.10	1.09	1.08
MeCl	1/06/89	15:51	56.0	21.2	76.0				1.2	48.0	26.8	42.0	100	68.0	5	54	74	60	500	1.10	1.09	1.08
MeCl	1/06/89	15:54	56.0	21.3	76.3					48.0	26.8	42.0	100	68.0	5				500	1.10	1.09	1.08
MeCl	1/06/89	16:00	56.0	21.5	76.7	45		200	2.5	47.0	26.8			69.0	5	54	73			1.10	1.09	1.08
MeCl	1/06/89	16:04	55.0	21.5	76.4	45		200	2.5	48.0	26.8			69.0	5	54	73			1.10	1.09	1.08
MeCl	1/06/89	16:09	56.0	21.6	76.2				2.5	48.0	26.8	84.0	50	69.0	5	54	72			1.10	1.09	1.08
MeCl	1/06/89	16:14	56.0	21.6	75.9				2.3	48.0	26.8	81.0	50	69.0	5	54	72			1.10	1.09	1.08
MeCl	1/06/89	16:18	56.0	21.3	76.0				1.6	47.5	26.8	80.0	50	69.0	5	54	72			1.10	1.09	1.08
MeCl	1/06/89	16:22	56.0	21.7	76.7				1.5	47.5	26.8	77.0	50	69.0	5	54	75			1.10	1.09	1.08
MeCl	1/06/89	16:28	56.0	21.5	76.6				2.3	47.5	26.8	76.0	50							1.10	1.09	1.08
MeCl	1/06/89	16:40	55.5	21.3	76.0				1.6	48.0	26.8			69.0	5	54	60			1.10	1.09	1.08
MeCl	1/06/89	16:55	56.0	21.5	76.6	76		100	2.3	47.5	26.8			69.0	5	55				1.10	1.09	1.08
MeCl	1/06/89	17:05	56.0	21.3	76.4	55		500	2.0	48.0	26.8			69.0	40	52	63			1.10	1.09	1.08
MeCl	1/06/89	17:10	56.0	21.6	77.0	55		2000	1.6	48.0	26.8			69.0	85	45	60			1.19	1.12	1.13
MeCl	1/06/89	17:15	56.0	21.5	76.7	64		2000	2.1	48.0	26.8			69.0	80	43	63			1.19	1.12	1.13
MeCl	1/06/89	17:20	56.0	21.5	76.5	68		2000	2.5	48.0	26.8			69.0	78	40	63			1.19	1.12	1.13
MeCl	1/06/89	17:25	56.0	21.7	76.3	65		2000	2.5	47.5	26.8			70.0	80	40	62			1.19	1.12	1.13

TABLE A-1. (CONTINUED)

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Solvent	Date	Time	InFlo	InPr	InTmp	I A	InAttn	OutPres	OutFlo	Vac	O.Area	OutAttn	SatTmp	SatFlo	SatPres	N2 Sup	Prm Ar	Prm Attn	I.Rf	O.Rf	PrmRf
MeCl	1/06/89	17:30	56.0	21.6	76.2	69	2000	1.5	48.0	26.8			69.0	79	38	61	98	5000	1.19	1.12	1.13
MeCl	1/06/89	17:35	56.0	21.5	76.7	71	2000	2.0	48.0	26.8			69.0	79	38	68	98	5000	1.19	1.12	1.13
MeCl	1/06/89	17:38	56.0	21.2	77.1	72	2000	1.6	48.0	26.8			69.0	79	38	72	98	5000	1.19	1.12	1.13
MeCl	1/06/89	17:46	56.0	21.8	76.5			2.4	49.0	26.8	63.0	1000	69.0	78	38	69	98	5000	1.19	1.12	1.13
MeCl	1/06/89	17:51	56.0	21.7	76.2			2.1	48.0	26.8	62.0	1000	69.0	78	38	67	98	5000	1.19	1.12	1.13
MeCl	1/06/89	17:55	56 0	21.5	75.8	73	2000	1.6	48.0	26.8			69.0	78	38	67	98	5000	1.19	1.12	1.13
MeCl	1/06/89	18:00	56.0	21.4	75.9	75	2000	1.3	48.0	26.8			69.0	79	38	67	98	5000	1.19	1.12	1.13
MeCl	1/06/89	18:05	56.0	21.3	76.5	76	2000	2.0	48.0	26 8			69.0	79	38	70	98	5000	1.19	1.12	1.13
MeCl	1/06/89	18:10	56.0	21.5	76.9			2.0	48.0	26.8	66.0	1000	69.0	79	38	70	98	5000	1.19	1.12	1.13
MeCl	1/06/89	18:14	56.0	21.7	76.7			2.0	48.0	26.8	66.0	1000	69.0	79	38	65	98	5000	1.19	1.12	1.13
MeCl	1/06/89	18:18	56.0	21.3	76.2	77	2000	1.9	48.0	26.8			69.0	79	38	69	98	5000	1.19	1.12	1.13
MeCl	1/06/89	18:22	56.0	21.2	76.2	78	2000	1.9	48.0	26.8			69.0	79	38	69	98	5000	1.19	1.12	1.13
MeCl	1/06/89	18:25	56.0	21.3	76.1	76	2000	2.1	48.0	26.8			69.0	79	38	69	98	5000	1.19	1.12	1.13
MeCl	1/06/89	18:30	56.0	21.7	76.9			2.0	48.0	26.8			69.0	79	38	63	98	5000	1.19	1.12	1.13
MeCl	1/06/89	18:35	56.0	21.1	76.8			1.3	48.0	26.8			69.0	79	38	71	98	5000	1.19	1.12	1.13
MeCl	1/06/89	18:40	56.0	21.7	76.8	22	2000	1.7	48.0	26.8			69.0	79	38	72	98	5000	1.19	1.12	1.13
MeCl	1/06/89	18:45	56.0	21.6	76.3	78	2000	1.8	48.0	26.8			69.0	79	38		98	5000	1.19	1.12	1.13
MeCl	1/06/89	18:50	56.0	21.6	76.2	77	2000	1.9	48.0	26.8			69.0	79	38	72	98	5000	1.19	1.12	1.13
MeOH	1/13/89	12:30	56.0	22.0	66.5	23	50	2.5	48.0	26.8			88.0			75					
MeOH	1/13/89	12:40	55.5	22.2	67.1			3.1	48.0	26.8	21.0	50	88.0			75					
MeOH	1/13/89	12:50	56.0	21.3	67.8	16	50	1.0	48.0	26.8			88.0			75					
MeOH	1/13/89	13:00	55.5	21.5	68.4			1.9	48.0	26.8	15.0	50	92.0		26	74					
MeOH	1/13/89	13:21	56.0	21.3	69.5	30	50	.1	48.5	26.8			88.0		38	63					
MeOH	1/13/89	13:35	56.0	21.7	70.0	51	50	1.2	48.5	26.8			89.0	16	38	72					
MeOH	1/13/89	13:50	56.0	22.0	70.4	71	100	1.5	48.5	26.8			90.0	30	37	70			.83	.60	1.01
MeOH	1/13/89	14:00	56.0	21.4	70.6	75	100	1.0	48.5	26.8			90.0	31	37	64	60	500	.83	.60	1.01
MeOH	1/13/89	14:10	56.0	21.6	70.6	71	100	1.5	48.5	26.8			89.0	24	34	71	60	500	.83	.60	1.01
MeOH	1/13/89	14:22	56.0	21.4	70.6			.9	48.5	26.8	81.0	50	90.0	24	38	73	60	500	.83	.60	1.01
MeOH	1/13/89	14:32	56.0	21.3	7.1			1.5	48.5	26.8	70.0	50	90.0	23	38	72	60	500	.83	.60	1.01
MeOH	1/13/89	14:40	56.0	21.3	71.0			1.0	48.5	26.8	67.5	50	90.0	23	38	71	60	500	.83	.60	1.01
MeOH	1/13/89	14:50	56.0	21.2	71.0			1.0	48.5	26.8	65.0	50	90.0	22	38	68	60	500	.83	.60	1.01
MeOH	1/13/89	15:05	56.0	21.1	71.1	64	100	.5	48.5	26.8			90.0	23	38	72	60	500	.83	.60	1.01
MeOH	1/13/89	15:20	56.0	21.6	71.2	70	100	1.2	48.5	26.8			90.0	22	38	68	60	500	.83	.60	1.01
MeOH	1/13/89	15:30	56.0	21.9	71.2			1.2	48.5	26.8			90.0	23	38	66	60	500	.83	.60	1.01
MeOH	1/13/89	15:45	56.0	21.2	71.6			1.2	48.5	26.8	81.0	50	90.0	22	37	70	60	500	.83	.60	1.01
MeOH	1/13/89	15:55	56.0	21.4	71.6			1.4	48.5	26.8	68.5	50	90.0	23	37	72	60	500	.83	.60	1.01
MeOH	1/13/89	16:10	56.0	21.6	71.7			1.5	48.5	26.8	64.0	50	90.0	23	38	69	60	500	.83	.60	1.01
MeOH	1/13/89	16:30	56.0	21.9	72.0	70	100	1.5	48.5	26.8			89.0	22	38	70	60	500	.83	.60	1.01
MeOH	1/13/89	16:48	56.0	21.8	72.1	47	500	2.2	48.5	26.8			99.0	72	36	66					
MeOH	1/13/89	17:00	56.0	21.3	72.0	68	500	1.2	48.5	26.8			109.0	72	36	68	60	500	1.01	.93	1.00
MeOH	1/13/89	17:25	56.0	21.4	72.0	76	500	1.0	48.5	26.8			111.0	72	35	69	60	500	1.01	.93	1.00
MeOH	1/13/89	17:45	56.0	21.3	71.6			.5	48.5	26.8	89.0	200	111.0	71	35	59	60	500	1.01	.93	1.00
MeOH	1/13/89	18:05	56.0	21.3	71.8	76	500	2.0	48.5	26.8			111.0	70	35	62	60	500	1.01	.93	1.00
MeOH	1/13/89	18:25	56.0	21.5	71.8			1.8	49.0	26.8	90.0	200	111.0	70	35	60	60	500	1.01	.93	1.00
MeOH	1/13/89	18:48	56.0	21.1	71.4			.3	48.5	26.8	89.0	200	111.0	70	35	60			1.00	1.00	1.00
MeOH	1/16/89	13:17	54.0	20.9	70.5	4	2000	2.0	47.0	26.8			162.0	20	38	55			1.00	1.00	1.00
MeOH	1/16/89	13:30	55.0	22.1	71.4	18	2000	2.5	47.0	26.8			168.0	40	45	56			1.00	1.00	1.00
MeOH	1/16/89	14:17	55.5	20.8	77.7	84	2000	1.0	48.0	26.8			167.0	70	60	67			1.00	1.00	1.00
MeOH	1/16/89	14:30	55.5	20.9	72.4	54	2000	2.0	48.0	26.8			166.0	70	57	59			1.00	1.00	1.00

TABLE A-1. (CONTINUED)

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Solvent	Date	Time	InFlo	InPr	InTmp	I	A	InAttn	OutPres	OutFlo	Vac	O.Area	OutAttn	SatTmp	SatFlo	SatPres	N2 Sup	Prm Ar	Prm Attn	I.Rf	O.Rf	PrmRf
MeOH	1/16/89	14:39	56.0	21.0	72.6	53		2000	1.5	48.0	26.8			165.0	70	54	53			1.00	1.00	1.00
MeOH	1/16/89	14:55	56.0	21.3	72.6				1.0	48.5	26.8	51.0	1000	165.0	70	66	58			1.00	1.00	1.00
MeOH	1/16/89	15:00	56.0	22.5	73.0				3.0	48.0	26.8	54.0	1000	165.0	70	67	59			1.00	1.00	1.00
MeOH	1/16/89	15:05	56.0	21.6	73.2				1.2	48.0	26.8	55.0	1000	165.0	70	66	56			1.00	1.00	1.00
MeOH	1/16/89	15:15	55.5	21.7	752.8	61		2000	1.4	48.0	26.8			165.0	70	66	62			1.00	1.00	1.00
MeOH	1/16/89	15:20	56.0	21.2	73.0	63		2000	2.0	48.0	26.8			165.0	70	66	72			1.00	1.00	1.00
MeOH	1/16/89	15:30	56.0	21.2	73.4	61		2000	1.5	48.0	26.8			165.0	70	66	53			1.00	1.00	1.00
MeOH	1/16/89	15:40	56.0	21.4	73.5				1.5	48.0	26.8	63.0	1000	165.0	70	65	58			1.00	1.00	1.00
MeOH	1/16/89	15:45	56.0	21.3	73.6				1.2	48.0	26.8	62.0	1000	165.0	70	65	63			1.00	1.00	1.00
MeOH	1/16/89	15:50	55.5	21.2	73.7	58		2000	.7	48.0	26.8			165.0	70	65	71			1.00	1.00	1.00
MeOH	1/16/89	15:55	56.0	21.7	73.8	61		2000	1.5	48.0	26.8			165.0	70	65	56			1.00	1.00	1.00
MeOH	1/16/89	16:00	55.5	21.3	73.8	62		2000	1.5	48.0	26.8			165.0	70	65	65			1.00	1.00	1.00
MeOH	1/16/89	16:06	56.0	21.7	74.0	61		2000	1.6	48.0	26.8			165.0	70	65	49			1.00	1.00	1.00
MeOH	1/16/89	16:18	56.0	21.8	73.3				1.4	48.0	26.8			165.0	70	65				1.00	1.00	1.00
MeOH	1/16/89	16:25	56.0	21.4	73.4	62		2000	1.9	48.0	26.8			165.0	70	65	56			1.00	1.00	1.00
MeOH	1/16/89	16:28	56 J	21.2	73.5	61		2000	1.2	48.0	26.8			165.0	70	65	58			1.00	1.00	1.00
MeOH	1/16/89	16:35	56.0	21.4	73.6				1.9	48.0	26.8	60.0	1000	165.0	70	65	62	40	10000	1.00	1.00	1.00
MeOH	1/16/89	16:40	55.5	21.3	73.8				1.2	48.0	26.8	58.0	1000	165.0	70	65	65	40	10000	1.00	1.00	1.00
MeOH	1/16/89	16:45	56.0	21.5	74.0				1.5	48.0	26.8	58.0	1000	165.0	70	65	53	40	10000	1.00	1.00	1.00
MeOH	1/16/89	16:50	56.0	22.0	73.4	60		2000	1.2	48.0	26.8			165.0	70	65	53	40	10000	1.00	1.00	1.00
MeOH	1/16/89	16:55	55.0	21.2	73.6	58		2000	1.0	48.0	26.8			165.0	70	65	57		10000	1.00	1.00	1.00
MeOH	1/16/89	17:00	56.0	21.1	74.0	59		2000	1.0	48.0	26.8			165.0	70	65				1.00	1.00	1.00
Toluene	1/19/89	16:33	55.5	20.1	69.5	5		500	1.0	48.0	26.8			68.0		24	76			1.11	1.03	1.14
Toluene	1/19/89	16:58	56.0	21.9	70.8	49		500	2.0	48.0	26.8			68.0	29	36	58			1.11	1.03	1.14
Toluene	1/19/89	17:03	56.0	21.6	70.8	50		500	2.0	48.0	26.8			68.0	31	36	60	40	2000	1.11	1.03	1.14
Toluene	1/19/89	17:15	56.0	21.3	71.8				1.0	48.0	26.8	58.5	200	68.0	31	37	65	40	2000	1.11	1.03	1.14
Toluene	1/19/89	17:20	56.0	21.5	71.9				2.0	48.0	26.8	56.0	200	69.0	32	37	73	40	2000	1.11	1.03	1.14
Toluene	1/19/89	17:25	56.0	21.1	71.7				.8	48.0	26.8	55.0	200	69.0	31	37	72	40	2000	1.11	1.03	1.14
Toluene	1/19/89	17:37	56.0	21.1	72.4	51		500	1.2	48.0	26.8			68.0	31	37	74	40	2000	1.11	1.03	1.14
Toluene	1/19/89	17:43	56.0	21.1	72.8	53		500	.6	48.0	26.8			69.0	31	37	65	40	2000	1.11	1.03	1.14
Toluene	1/19/89	17:48	56.0	21.5	73.0	52		500	1.9	48.0	26.8			69.0	31	37	69	40	2000	1.11	1.03	1.14
Toluene	1/19/89	18:06	56.0	21.0	73.0				1.5	48.0	26.8	57.0	200	69.0	30	37	72	40	2000	1.11	1.03	1.14
Toluene	1/19/89	18:11	56.0	21.3	73.2				1.0	48.0	26.8	56.0	200	69.0	31	37	72	40	2000	1.11	1.03	1.14
Toluene	1/19/89	18:16	56.0	21.5	73.3				1.2	48.0	26.8	56.5	200	69.0	31	37	66	40	2000	1.11	1.03	1.14
Toluene	1/19/89	18:25	56.0	21.3	73.2	51		500	1.9	48.0	26.8			69.0	30	37	73	40	2000	1.11	1.03	1.14
Toluene	1/19/89	18:30	55.5	21.3	72.9	52		500	2.5	48.0	26.8			69.0	25	36	73	40	2000	1.11	1.03	1.14
Toluene	1/19/89	18:40	56.0	21.2	73.2				2.4	48.0	26.8			69.0	26	37	65	40	2000	1.11	1.03	1.14
Toluene	1/19/89	18:45	55.5	21.6	72.9				1.6	48.0	26.8			69.0	26	37	62	40	2000	1.11	1.03	1.14
Toluene	1/19/89	18:59	56.0	21.0	73.2	51		500	4.0	48.0	26.8			69.0	26	36	71	40	2000	1.11	1.03	1.14
Toluene	1/19/89	19:04	56.0	21.0	73.0	52		500	2.0	48.0	26.8			69.0	27	36	73			1.11	1.03	1.14
Toluene	1/19/89	19:15	56.0	21.2	73.6	45		2000	1.0	48.5	26.8			76.0	67	32	67			1.15	1.14	1.07
Toluene	1/19/89	19:20	56.0	21.4	73.7	49		2000	1.5	48.5	26.8			84.0	65	32	61			1.15	1.14	1.07
Toluene	1/19/89	19:25	56.0	21.8	74.0	49		2000	1.1	48.5	26.8			89.0	66	32	62			1.15	1.14	1.07
Toluene	1/19/89	19:30	56.0	21.5	74.1	50		2000	1.9	48.5	26.8			90.0	65	32	67	60	5000	1.15	1.14	1.07
Toluene	1/19/89	19:40	56.0	21.5	73.2				1.5	48.5	26.8	41.5	1000	90.0	65	32	65	60	5000	1.15	1.14	1.07
Toluene	1/19/89	19:45	56.0	21.4	73.4				1.5	48.5	26.8	41.0	1000	90.0	66	32	67	60	5000	1.15	1.14	1.07
Toluene	1/19/89	19:50	56.0	21.3	73.5				1.5	48.5	26.8	42.0	1000	90.0	67	32	67	60	5000	1.15	1.14	1.07
Toluene	1/19/89	20:00	56.0	21.3	73.6	49		2000	1.7	48.5	26.8			90.0	67	32	72	60	5000	1.15	1.14	1.07
Toluene	1/19/89	20:05	56.0	21.2	73.2	51		2000	1.5	48.5	26.8			90.0	75	33	67	60	5000	1.15	1.14	1.07

TABLE A-1. (CONTINUED)

2/09/89 chkdat Page 6

Date	Time	InFlo	InPr	InTmp	I A	InAttn	OutPres	OutFlo	Vac	O.Area	OutAttn	SatTmp	SatFlo	SatPres	N2 Sup	Prm Ar	Prm Attn	I.Rf	O.Rf	PrmRf	
Toluene	1/19/89	20:10	55.5	21.5	73.4	51	2000	1.6	48.0	26.8			90.0	75	33	66	60	5000	1.15	1.14	1.07
Toluene	1/19/89	20:17	55.5	21.3	73.9			1.7	48.0	26.8			90.0	75	33	64	60	5000	1.15	1.14	1.07
Toluene	1/19/89	20:23	55.5	21.3	73.8			2.0	48.0	26.8			90.0	74	33	64	60	5000	1.15	1.14	1.07
Toluene	1/19/89	20:40	56.0	21.4	73.6	51	2000	2.2	48.0	26.8			90.0	68	33	66	60	5000	1.15	1.14	1.07
Toluene	1/19/89	20:45	55.5	21.3	74.0	53	2000	1.1	48.0	26.8			90.0	68	33	66	60	5000	1.15	1.14	1.07
Toluene	1/19/89	20:50	55.5	21.2	73.9	53	2000	2.4	48.0	26.8			90.0	70	33	68	60	5000	1.15	1.14	1.07
Toluene	1/19/89	20:55	55.5	21.3	73.2	54	2000	1.3	48.0	26.8			90.0	72	33	66	60	5000	1.15	1.14	1.07
Toluene	1/19/89	21:05	56.0	21.3	73.6			1.5	48.0	26.8	47.0	1000	90.0	72	33	68	60	5000	1.15	1.14	1.07
Toluene	1/19/89	21:10	56.0	21.3	73.7			1.5	48.5	26.8	48.0	1000	90.0	72	33	70	60	5000	1.15	1.14	1.07
Toluene	1/19/89	21:15	56.0	21.3	74.0			1.5	48.0	26.8	47.5	1000	90.0	72	33	71	60	5000	1.15	1.14	1.07
Toluene	1/19/89	21:20	56.0	21.5	73.4	50	2000	2.0	48.5	26.8			90.0	72	33	68	60	5000	1.15	1.14	1.07
Toluene	1/19/89	21:25	56.0	21.5	73.3	52	2000	2.0	48.0	26.8			90.0	73	33	67	60	5000	1.15	1.14	1.07
Toluene	1/19/89	21:30	56.0	21.5	73.5	53	2000	1.4	48.0	26.8			90.0	70	33	67	60	5000	1.15	1.14	1.07
Toluene	1/19/89	21:35	56.0	21.2	73.7	53	2000	.9	48.0	26.8			90.0	72	33	67	60	5000	1.15	1.14	1.07
Toluene	1/19/89	21:40	56.0	21.8	74.0			1.2	48.5	26.8			90.0	72	33	60	60	5000	1.15	1.14	1.07
Toluene	1/19/89	21:50	56.0	21.7	73.2			1.4	48.5	26.8			90.0	70	33	61	60	5000	1.15	1.14	1.07

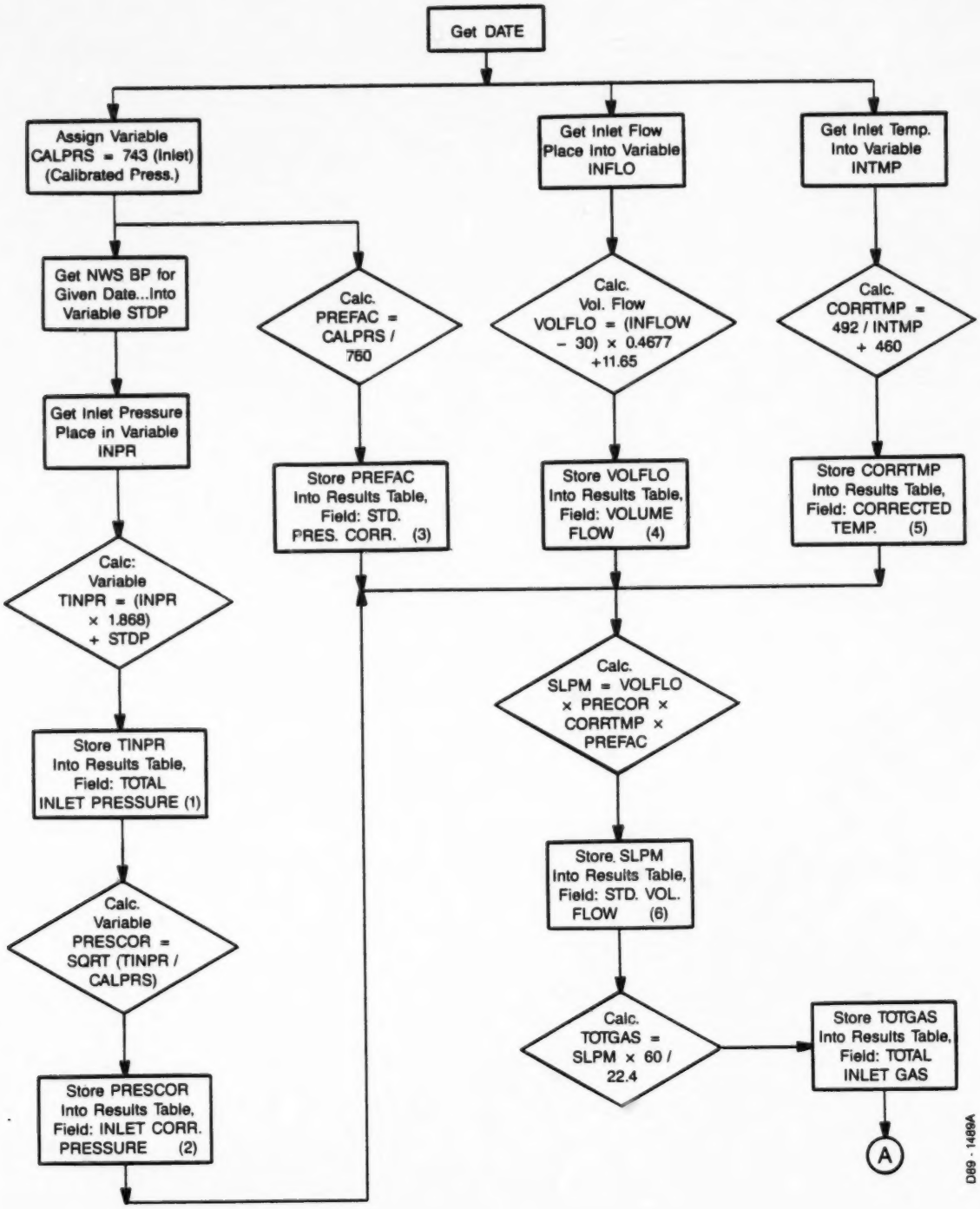
Get DATE
Assign Variable CALPRS = 743 (Inlet) (Calibrated Press.)
Get Inlet Flow Place Into Variable INFLO
Get Inlet Temp. Into Variable INTMP
Get NWS BP for Given Date...Into Variable STDP
Calc. PREFAC = CALPRS / 760
Calc. Vol. Flow VOLFLO = (INFLOW - 30) x 0.4677 +11.65
Calc. CORRTMP = 492 / INTMP + 460
Get Inlet Pressure Place in Variable INPR
Store PREFAC Into Results Table, Field: STD. PRES. CORR. (3)
Store VOLFLO Into Results Table, Field: VOLUME FLOW (4)
Store CORRTMP Into Results Table, Field: CORRECTED TEMP. (5)
Calc: Variable TINPR = (INPR x 1.868) + STDP
Calc. SLPM = VOLFLO x PRECOR x CORRTMP x PREFAC
Store TINPR Into Results Table, Field: TOTAL INLET PRESSURE (1)
Store SLPM Into Results Table, Field: STD. VOL. FLOW (6)
Calc. Variable PRESCOR = SQRT (TINPR / CALPRS)
Calc. TOTGAS = SLPM x 60 / 22.4
Store TOTGAS Into Results Table, Field: TOTAL INLET GAS
Store PRESCOR Into Results Table, Field: INLET CORR. PRESSURE (2)
A
D69 - 1469A

Figure A-1. Example Calculations

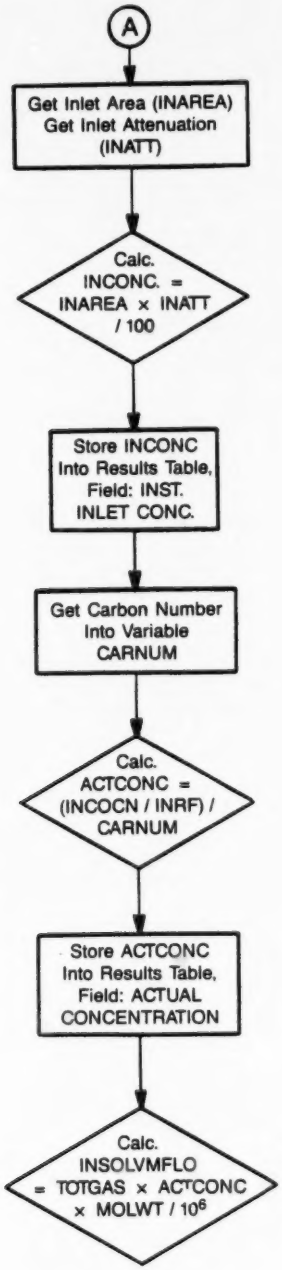

Figure A-1. (Continued)

(1) Total Inlet Pressure = Inlet Pressure X 1.868 + Local Std. Bar. Pressure

(2) Inlet Corrected Pressure = $\sqrt{\frac{\text{Total Inlet Pressure}}{\text{Instrument Calibrated Pressure}}}$

(3) Std. Corrected Pressure = $\frac{\text{Instrument Calibrated Pressure}}{\text{Standard Pressure}}$

(4) Uncorrected Vol. Flow = (Inlet Flow - 30) X 0.4677 + 11.65

(5) Corrected Temperature = 492/(Inlet Temp + 460)

(6) Std. Volume Flow = Inlet Corr Pres X Std Corr Pres X Vol Flow X Corr Temp
(2) (3) (4) (5)

(7) Total Inlet Gas = Std Vol. Flow X 60/22.4
(6)

(8) Instrument Inlet Concentration = Inlet Area * Inlet Attenuation/100

(9) Actual (Solvent) Concentration = $\left(\frac{\text{Inst. Inlet Conc.}}{\text{Inlet Rf}}\right)$ / Carbon Number

(10) Inlet Solvent Mass Flow =

Total Inlet Gas X Actual Concentration X Molecular Wt. / 10^6
(7) (9)

Figure A-1. (Continued)

APPENDIX B
QC RESULTS

CALIBRATIONS

Temperature

The temperature of the inlet gas and the liquid in the saturator were measured with calibrated thermocouples. Each thermocouple was checked against an NBS-traceable calibration thermometer by immersing them both in dewar flasks containing water at various temperatures. The results of the calibration are shown in Table B-1.

Pressure

Pressure was measured at several locations in the apparatus; these consisted of inlet gas pressure to the membrane, outlet gas pressure (downstream of the outlet rotameter), vacuum level in the permeate stream, pressure in the saturator, and nitrogen make-up supply pressure (downstream of the rotameter). Each pressure gauge was checked against a reference manometer or gauge by connecting both gauges to sources at various pressures. The results of the calibrations are shown in Table B-2.

TABLE B-1. TEMPERATURE SENSOR CALIBRATION

	Indicated Temp. (°F)	Actual Temp. (°F)	Date
A. Saturator Thermocouple	63	66.7	1-9-89
	77	80.0	
	96	99.1	
	115	118.5	
	132	135.6	
	146	149.0	
	176	179.0	
	209	211.6	
B. Inlet Gas Thermocouple	68.2	67.8	1-9-89
	75.1	77.5	
	91.1	91.8	
	166.7	167.0	

TABLE B-2. PRESSURE GAUGE CALIBRATION

	Indicated Pressure (inches H_2O)	Reference Manometer (inches H_2O)	Date
A. Inlet Pressure	0.6	0.6	1-23-89
	12	12	
	18	18	
	24	24	
	30	30.4	
B. Outlet Pressure (Magnehelic)	5	5	1-23-89
	10	10	
	15	15	
	20	20	
	25	25.3	

	Indicated Pressure	Reference Gauge	Date
C. Nitrogen Supply Pressure (Bourdon Tube)	11.80 psig	12.00 psig	1-23-89
	8.70 psig	8.65 psig	
	0.55 psig	0.60 psig	
D. Vacuum Gauge (Bourdon Tube)	29.0 in Hg	28.91 in Hg	1-23-89
	24.4 in Hg	24.23 in Hg	
	0.0 in Hg	0.00 in Hg	

Flow

Flow was measured at several locations in the apparatus; these consisted of inlet gas flow to the membrane, outlet gas flow from the membrane, saturator flow to the recirculating loop, and nitrogen make-up flow. Each of these flowrates were measured using calibrated rotameters. Each rotameter was calibrated with nitrogen at a known discharge pressure using a Hastings-Raydist Flow Calibrator ("soap bubble flow meter"). Results of the calibrations are shown in Table B-3.

Total Hydrocarbon (THC) Analyzer

The Byron 401 THC Analyzer was used to measure solvent concentration in the inlet, outlet, and permeate streams of the membrane. The analyzer itself was checked weekly for proper operation by performing a multipoint calibration with propane in nitrogen. Table B-4 presents the actual propane concentrations, instrument concentrations, and average response factors for the weekly calibrations. Also shown are the linear correlation coefficients ("r") for the relationship of actual concentrations versus instrument readings. All the propane calibrations exceeded the minimum acceptance criteria of $r \geq 0.995$.

Trap Recovery

The gravimetric cold trap was tested for its ability to recover at least 95% of the incoming solvent vapor. The results for these tests are presented in Table B-6. The results were disappointing, in that the goal of 95% recovery was not met. It appeared that even with two traps in series, a small amount of the solvent was still escaping (8%). With no alternative sampling method available at the early stages of the testing, it seemed possible that using liquid nitrogen only in the first (gravimetric) trap might improve the recovery. This procedure was used from 12/15/88 until the permeate sampling pumps were added on 1/6/89.

TABLE B-3. FLOWMETER CALIBRATIONS

	Setting	Flowrate	Barometric Pressure	Date
A. Inlet Flow Rotameter	30	11.65 L/min	743 mmHg	11-11-88
	60	25.68		
	90	40.81		
	120	57.10		
B. Outlet Flow Rotameter	20	7.28 L/min	735 mmHg	11-15-88
	40	15.73		
	60	25.42		
	80	35.77		
	110	53.19		
C. Nitrogen Supply Rotameter	20	0.929 L/min	743 mmHg	11-11-88
	60	3.554		
	100	6.108		
	150	9.255		
D. Saturator Flow Rotameter	10	20.09 mL/min	742 mmHg	1-25-89
	30	36.91		
	50	67.38		
	70	113.1		
	90	172.6		

TABLE B-4. WEEKLY PROPANE MULTIPOINT CALIBRATIONS

Date	Actual Conc. (ppmvC)	Instrument Conc. (ppmvC)	Correlation Coefficient	Average RF
12-14-88	43	63	r=0.9991	1.463
	111	166		
	415	645		
	929	1330		
	2302	3400		
	4547	6750		
	12277	16400		
12-15-88	0	7.5	r=.9991	1.444
	111	167		
	929	1350		
	4548	6750		
	12276	14600		
12-19-88	0	6	r=.9991	1.470
	111	175		
	929	1360		
	4548	6800		
	12276	16500		
1-3-89	0	5	r=.9994	1.441
	111	166		
	929	1340		
	4548	6700		
	12276	16600		
1-9-89	0	3	r=.9994	1.443
	111	168		
	929	1340		
	4548	6700		
	12276	16500		
1-16-89	0	6.5	r=.9991	1.451
	111	168		
	929	1350		
	4548	6800		
	12276	16500		

TABLE B-5. GRAVIMETRIC TRAP RECOVERY TEST RESULTS

Date	Time	Solvent Added	Primary Trap	Back-up Trap	% Recovery	% Closure
11-23-88	14:15	6.0 ml MeCl	6.36 g	-	80.4%	-
11-23-88	14:48	7.0 ml MeCl	7.64 g	1.0 ml MeCl	82.7%	97.1%

TEST CONDITIONS:

1. Syringe injection time varied from 1 to 2.5 min.
2. Vacuum level in line of 24 in Hg.
3. Primary trap cold bath at -150°C.

Solvent Response Factors

The Byron 401 THC Analyzer was checked daily with a multipoint calibration with standard gases of the solvent. Results of these calibrations are shown in Table B-6.

The response factors (RFs) are calculated using the formula shown below:

$$RF = \frac{\text{Instrument Concentration (ppmvC)}}{\text{Actual Concentration (ppmvC)}}$$

where, Instrument Concentration = (Area)*(Attenuation)/100

Actual Concentration = (Solvent Standard Concentration)*(No. of Carbon Atoms)

Permeate THC samples

As an improvement over the gravimetric cold trap to sample the permeate THC concentration, additional sample pumps were connected in series to extract a sample from the permeate stream. The sampling technique was checked by metering one of the solvent standard gases into the permeate line and checking the response of the THC analyzer. The vacuum level in the line was maintained at about 27 in Hg to simulate experimental conditions. As shown in Table B-7, the sample pumps appeared to cause some dilution, although it was difficult to determine how much.

TABLE B-6. DAILY SOLVENT MULTIPOINT CALIBRATION DATA

Solvent	Date	Concentration Range, ppmv (ppmvC)	Average RF	Correlation Coefficient r
Hexane	12-14-88	0-141 ppmv (0-848 ppmvC)	1.286	0.9997
	12-15-88	0-2000 ppmv (0-12000 ppmvC)	1.310	0.9984
	1-20-89	0-2000 ppmv (0-12000 ppmvC)	1.307	0.9986
MEK*	12-16-88	0-1980 ppmv (0-7920 ppmvC)	0.967	0.9993
	12-19-88	0-1980 ppmv (0-7920 ppmvC)	0.986	0.9994
	1-3-89	0-1980 ppmv (0-7920 ppmvC)	0.955	0.9989
MeCl	1-4-89	0-2000 ppmv (0-2000 ppmvC)	1.220	0.9989
	1-5-89	0-141 ppmv (0-141 ppmvC)	1.096	0.9990
	1-6-89	0-2000 ppmv (0-2000 ppmvC)	1.114	0.9998
CFC-113	1-9-89	0-1980 ppmv (0-3960 ppmvC)	1.031	0.9997
	1-10-89	0-1980 ppmv (0-3960 ppmvC)	1.046	0.9996
MeOH*	1-13-89	0-1000 ppmv (0-1000 ppmvC)	0.834	0.9998
	1-16-89	0-1000 ppmv (0-1000 ppmvC)	0.981	0.9998
Toluene	1-19-89	0-500 ppmv (0-3500 ppmvC)	1.101	0.9993

* - RF decreases at low concentrations.

TABLE B-7. DIRECT PERMEATE SAMPLING TEST

Date	Sample	BYRON 401 Response	% Recovery
1-13-89	1*	4100 ppmvC	100%
	2	3791	92.5%
	3	3600	87.8%

*1 - Calibration gas connected directly to analyzer sample inlet.

2 - Calibration gas connected to inlet of sample pump.

3 - Calibration gas connected to blanked-off permeate side with vacuum at 27 in Hg.

APPENDIX C
LABORATORY SYSTEMS AUDIT REPORT

RADIAN
CORPORATION

AUDIT REPORT 259-044-01-04

15 March 1989

TO: Distribution
FROM: J.M. Youngerman
SUBJECT: January 1989 ARB Membrane Bench Scale Laboratory Systems Audit
PROJECT: California Air Resources Board/U.S. EPA Membrane Test Study

1.0 INTRODUCTION

A Technical Systems Audit (TSA) of the Air Resources Board (ARB) Bench Scale Laboratory was conducted by Jean Youngerman on January 19, 1989. This audit covered general laboratory operations and specific requirements for the support of the quality assurance (QA) effort for the ARB Membrane Test Study. This systems audit was conducted to determine the extent to which the QAPP is currently followed.

Technical Systems Audits are conducted to evaluate the adequacy of the measurement system to provide data of known quantity which are sufficient, in terms of quality and quantity, to meet the program objectives.

The current audit focused on the following areas:

- Condition of the facilities and equipment;
- Consistency of current practices with documented procedures presented in the ARB Membrane QAPP;
- Calibration procedures and documentation;
- Completeness of data forms and data reduction procedures;

- Recordkeeping (data filing and archiving procedures); and
- Compliance with the ARB Membrane Test Study QAPP quality control requirements.

Specific activities underway during the audit included the final analyses of one of the test gases for the ARB Membrane Test Study. The results of the current audit are documented in the checklists presented as Attachment A of this report.

2.0 RECOMMENDATIONS FOR CORRECTIVE ACTION

No formal Recommendations for Corrective Action (RCAs) are being issued at this time since the program was in nearly complete at the time of the audit.

3.0 DISCUSSION

A checklist was used as a guide in conducting this systems audit. The completed checklist is presented as Attachment A of this report. Each checklist was designed to document the status of those elements which are critical to production of defensible data of known quality. In addition, project-specific elements required by the ARB Membrane Test Study QAPP were reviewed and documented.

3.1 Facilities and Equipment

The bench-scale laboratory of the ARB Membrane Test Study is maintained at a reasonable level of cleanliness and order. Instrumentation in use for the program is mechanically sound. This instrumentation meets the specification presented in the ARB Membrane Test Study QAPP and should be technically suitable for the intended purpose. At the time of the audit the instrumentation was operational. No problems were noted.

During the course of the study, it was determined that the cold trap procedure described in the QAPP would not be adequate for the needs of the project. Therefore, a direct permeate sampling method was used to obtain the sample from the permeate.

3.2 Consistency with Documented Procedures

Documented operating procedures for the ARB Membrane Test Study are presented in the ARB Membrane Test Study QAPP. The QAPP includes procedures to be followed for sample handling, sample analysis and data reporting. The QAPP requires that a control standard be analyzed in duplicate after the drift check analyses. The control standard was not purchased. Other quality control procedures were followed as specified in the QAPP.

3.3 Calibration Procedures and Documentation

The THC analyzer calibration was performed according to the QAPP. At the time of the audit, none of the calculations for the propane multipoint calibration had been performed. Therefore, the analyst was not sure if the multipoint calibration curve met the stated acceptance criteria of $r \geq 0.995$. The auditor explained to the analyst the reasoning for calculating the linear regression of the calibration curve after the curve is run. It was informally recommended that these calculations be made immediately. After the audit, the analyst confirmed the linearity of the calibration curve. The standard multipoint calibration was performed daily; therefore, a drift check was not required.

At the time of the audit, the following calibrations needed to be completed and documented:

- Saturator rotameter;
- Pressure gauges; and
- Vacuum gauges.

The analyst planned on checking the calibration of these instruments as soon as the testing was complete.

3.4 Completeness of Data Forms and Data Reduction Procedures

The data forms presented in the QAPP had been modified to include more information. The forms were filled out correctly. The laboratory calculations required for this program had been performed. However, no sample calculations had been given. It was informally recommended that sample calculations be added to the logbook. The calculations had been reviewed by a supervisor, but there were no initials to evidence this review. No data reduction, using a database, had begun at the time of this audit.

3.5 Recordkeeping

At the time of this audit, recordkeeping consisted of some chromatograms filed in manila folders and a project logbook. The chromatograms contained all the necessary information including: dates, initials, and amounts.

However, entries in the logbook were not signed. Corrections were made by crossing out entire pages. These corrections were not dated, initialed, or explained. Although supervisor review of the notebook had occurred, it was not evidenced by initials.

3.6 Compliance with ARB Membrane Test Study QAPP Quality Control Requirements

At the time of this audit, not all of the QC protocol specified in the QAPP was in use. Although calibration curves were in use, they had not been calculated, and no checks of instrument response consistency from day to day were employed. Because the standard gas multipoint curve was analyzed daily, a drift check was never run. The samples were analyzed

continuously at a constant temperature. This gives a measure of repeatability of the analyses. Blank samples were being analyzed. The data forms were complete, and the logbooks were in good shape.

4.0 STATUS OF PREVIOUS RECOMMENDATIONS

This audit report is the only one for this laboratory and the ARB Membrane Program. No formal Recommendations for Corrective Action are presented.

APPENDIX D
EXAMPLE DESIGN AND COST CALCULATIONS

TABLE D-1. EXAMPLE DESIGN CALCULATIONS

A. Carbon Adsorber Only

- Given: Inlet Flow = 2500 ACFM (2196 SCFM)
  Inlet Concentration = 1000 ppmv CFC-113

1. Calculate Equilibrium Carbon Loading: [use Calgon adsorption isotherm graph]

$$\text{Partial Pressure} = (y)(P_t) = \left(\frac{1000\text{ppmv}}{10^6}\right)(15.0\text{ psia})$$

$$= \underline{0.015\text{ psia}}$$

$$\text{Capacity at } 100°F = \underline{45\text{ wt\%}} \qquad \frac{0.45\text{ kg CFC-113}}{\text{kg carbon}}$$

2. Calculate Mass of Carbon Required: (From Ref. 10)

$$R_a = 1.556 \times 10^{-7}\,(MW)(C)(G)(t)/W_c$$

where: R_a = Total carbon required (lb)
MW = Molecular weight of solvent (CFC-113=187.4 lb/lb mol)
C = Concentration of solvent (1000 ppmv)
G = Vent stream flowrate (2196 SCFM)
t = Breakthrough time (use 2.5 hours as first guess)
W_c = Working capacity = (equilibrium capacity)/2
= (.45/2) = .225 lb CFC-113/lb carbon

$$R_a = (1.556\times10^{-7})(187.4)(1000)(2196)(2.5)/(.225) =$$

$\underline{711\text{ lb}}$ carbon required.

3. Calculate Desired Superficial Velocity: (From Ref. 10)

$$V_s = -.01576(C) + 101.579$$

where: V_s = Superficial velocity (ft/min)
C = Concentration of solvent ($100<C<2{,}000$)

for C = 1000 ppmv, $V_s = -.01576(1000) + 101.579 = \underline{75\text{ ft/min}}$

(Continued)

TABLE D-1. (Continued)

4. Calculate Bed Area Normal to Flow: (From Ref. 10)

$A_o = (G)(460 + T)/(528V_s)$

where: A_o = Bed area normal to flow (ft^2)
G = Vent stream flowrate (2196 SCFM)
T = Gas inlet temp (100°F)
V_s = Superficial velocity (75 ft/min)

$A_o = (2196)(460 + 100)/(528 \cdot 75) =$ 31 ft^2

5. Calculate Bed Depth: (From Ref. 10)

$d = R_a/(A_o \cdot D)$

where: d = Depth of bed (ft);
R_a = Required amount of carbon (711 lb)
A_o = Bed area (31 ft^2)
D = Density of carbon (30 lb/ft^3)
$d = 711/(31 \cdot 30) =$ 0.76 ft

5a. If bed depth is less than 1.5 ft, the total carbon required should be recalculated using a bed depth of 1.5 ft:

$R_a = (d)(A_o \cdot D) = (1.5)(31 \cdot 30) =$ 1400 lb

5b. Recalculate the breakthrough time (no longer equal to 2.5 hours):

$t = R_a \cdot W_c/[1.556 \times 10^{-7}(MW)(C)(G)]$

$= 1400 \cdot (.225)/[1.556 \times 10^{-7}(187.4)(1000)(2196)]$

$=$ 4.9 hrs

B. Membrane Design with Carbon Adsorber (Using Vacuum Pump)

- Given: Membrane Selectivity (α^*) = 20
  Pressure Ratio (γ) = 0.10
  Removal Efficiency = 60%
  (Set by adjusting x and calculating F and y below)

(Continued)

TABLE D-1. (Continued)

6. Calculate Stage Cut (or "Fraction Permeated"): (From Ref. 2)

$F = 1 - (x/xf)^{[\alpha^*/(1-\gamma)(\alpha^*-1)]-1}$

where: F = Stage cut (molar flow, permeate/molar flow, inlet)
x_f = Concentration of solvent in feed (1000 ppmv or .001)
x = Concentration of solvent in residue (450 ppmv or .00045)
α^* = Selectivity (20)
γ = Pressure ratio (0.10)

$$F = 1 - \left(\frac{0.00045}{0.001}\right)^{[20/(1-0.1)(20-1)]-1}$$

$$= \underline{0.1266}$$

7. Calculate Permeate Concentration: (From Ref. 2)

$y = x_f/F\ [1 - (x/x_f)^{\alpha^*/(1-\gamma)(\alpha^*-1)}]$

where: y = Permeate concentration

$$y = \frac{0.001}{0.1266} = \left[1 - \left(\frac{0.0045}{0.001}\right)^{20/(1-0.1)(20-1)}\right]$$

= 0.004793 (4790 ppmv)

8. Calculate Dimensionless Area: (From Ref. 2)

$R_f = F/(1-\gamma)$

where: R_f = Dimensionless area: $(Q_b/d)PS/L_f$
F = Stage cut (0.1266)
γ = Pressure ratio (0.10)

$R_f = (.1266)/(1-.1) = \underline{0.1407}$

9. Calculate Membrane Area: (From Ref. 2)

$R_f = (Q_b/d)PS/L_f$

where: R_f = Dimensionless area (ft^2)
Q_b = Permeability of nitrogen (see below)
d = Membrane thickness (ft)

(Continued)

TABLE D-1. (Continued)

6. Calculate Stage Cut (or "Fraction Permeated"): (From Ref. 2)

$$F = 1 - (x/xf)^{[\alpha^*/(1-\gamma)(\alpha^*-1)]-1}$$

where: F = Stage cut (molar flow, permeate/molar flow, inlet)
x_f = Concentration of solvent in feed (1000 ppmv or .001)
x = Concentration of solvent in residue (450 ppmv or .00045)
α^* = Selectivity (20)
γ = Pressure ratio (0.10)

$$F = 1 - \left(\frac{0.00045}{0.001}\right)^{[20/(1-0.1)(20-1)]-1}$$

$$= \underline{0.1266}$$

7. Calculate Permeate Concentration: (From Ref. 2)

$$y = x_f/F\ [1 - (x/x_f)^{\alpha^*/(1-\gamma)(\alpha^*-1)}]$$

where: y = Permeate concentration

$$y = \frac{0.001}{0.1266} - \left[1 - \left(\frac{0.0045}{0.001}\right)^{20/(1-0.1)(20-1)}\right]$$

= 0.004793 (4790 ppmv)

8. Calculate Dimensionless Area: (From Ref. 2)

$$R_f = F/(1-\gamma)$$

where: R_f = Dimensionless area: $(Q_b/d)PS/L_f$
F = Stage cut (0.1266)
γ = Pressure ratio (0.10)

$R_f = (.1266)/(1-.1) = \underline{0.1407}$

9. Calculate Membrane Area: (From Ref. 2)

$$R_f = (Q_b/d)PS/L_f$$

where: R_f = Dimensionless area (ft^2)
Q_b = Permeability of nitrogen (see below)
d = Membrane thickness (ft)

(Continued)

TABLE D-1. (Continued)

P = Feed side pressure (psia)

S = Membrane area (ft^2)

L_f = Inlet molar flowrate (lb mol/hr)

Rearrange, using $\alpha^* = Q_a/Q_b$ or $Q_b = Q_a/\alpha^*$

$$S = \frac{R_f \cdot d \cdot L_f}{Q_b \cdot P} = \frac{\alpha^* R_f \cdot d \cdot L_f}{Q_a P}$$

where: Q_a = Permeability of solvent $\frac{\text{lb mol} \cdot \text{ft}}{\text{hr} \cdot \text{psi} \cdot \text{ft}^2}$

Q_a from Test Results:

$$\text{solvent flux}, \left(\frac{\text{gmol}}{\text{hr}}\right) \left(\frac{\text{membrane thickness, m}}{\text{membrane area, m}^2}\right) \left(\frac{1}{\text{partial pressure}}\right)$$

For CFC-113,

Q_a =

$$\left(.0477 \frac{\text{gmol}}{\text{hr}}\right)\left(\frac{3\text{x}10^{-6}\text{m}}{0.4\ \text{m}^2}\right)\left(\frac{1}{\frac{(1.077+0.54)}{2} - (0.307)}\right)\left(\frac{\text{lb mol}}{454\text{gmol}}\right)\left(\frac{760\ \text{mmHg}}{14.7\ \text{psia}}\right)\left(\frac{\text{m}}{3.281\ \text{ft}}\right)$$

$$= \underline{2.47 \times 10^{-8}}\ \frac{\text{lbmol} \cdot \text{ft}}{\text{ft}^2 \cdot \text{hr} \cdot \text{psi}}$$

Membrane area:

$$S = \frac{\alpha^* R_f \cdot d \cdot L_f}{Q_a \cdot P}$$

$$L_f = \left(2196 \frac{\text{std ft}^3}{\text{min}}\right)\left(\frac{\text{lb mol}}{359\ \text{ft}^3}\right)\left(\frac{60\ \text{min}}{\text{hr}}\right) = 367 \frac{\text{lb mol}}{\text{hr}}$$

$$S = \frac{(20)(0.1407)(3\text{x}10^{-6}\text{m})(3.281\ \text{ft/m})(367\ \text{lb mol/hr})}{2.47 \times 10^{-8}\ \frac{\text{lb mol} \cdot \text{ft}}{\text{ft}^2 \cdot \text{hr} \cdot \text{psi}}} \quad (15\ \text{psia})$$

$$= \underline{27{,}400}\ \text{ft}^2$$

10. Gas Flow to Carbon Adsorber:

$G' = F \cdot L_f/60$

(Continued)

TABLE D-1. (Continued)

where: G' = Molar flowrate to carbon adsorber (lb mol/min)

F = Stage cut

L_t = Inlet molar flowrate (lb mol/hr)

G' = (0.1266)(367)/60

= 0.774 lb mol/min

11. Actual Gas Volume to Vacuum Pump:

$$Q' = G' \left(359 \frac{ft^3}{lb\ mol}\right) \left(\frac{T + 460}{492}\right) \left(\frac{1}{\gamma}\right)$$

where: Q' = Actual cubic ft per min to vacuum pump (ACFM)

G' = Gas flow to adsorber (0.774 lb mol/min)

T = Inlet gas temperature (100°F)

γ = Pressure ratio (0.10)

Q' = (0.774)(359)(560/492)(10)

= 3160 ACFM

12. Required Horsepower for Vacuum Pump: (From Ref. 9)

$$W_s = \frac{(G')(MW)}{20} \frac{60\ min}{hr} \frac{[1/\gamma]^{0.286} - 1}{\eta}$$

where: W_s = Required horsepower of compression

G' = Gas flow to adsorber (.774 lb mol/min)

MW = Molecular weight of gas (29 lb/lb mol)

γ = Pressure ratio (0.10)

η = Adiabatic efficiency (use 0.36 for liquid ring pump)

$$W_s = \frac{(0.774)(29)}{20}(60)\frac{(1.958 - 1)}{.36}$$

W_s = 180 HP

(Continued)

TABLE D-1. (Continued)

13. Carbon Adsorber Design: (see steps 1-5 above)

- Calculate Equilibrium Loading (55 wt% CFC-113)
- SCFM to Adsorber (278 SCFM)
- Required Amount of Carbon (350 lb)
- Superficial Velocity (63 ft/min)
- Bed Area (4.7 ft^2)
- Bed Depth (2.5 ft)

C. Membrane Design with Carbon Adsorber (Using Compressor)

- Given: Membrane Selectivity (α^*) = 20

  Pressure Ratio (γ) = 0.2 (15/75)

  Removal Efficiency = 60%

14. Membrane Design: (see steps 6-13 above)

- Stage Cut (0.1981)
- Permeate Concentration (0.00304 or 3040 ppmv)
- Dimensionless Area (0.2476)
- Membrane Area (9646 ft^2)
- Compressor Work (334 HP) (Ref. 13)
- Gas Flow to Adsorber (1.212 lb mol/min)

15. Carbon Adsorber Design: (see steps 1-5 above)

- Calculate Equilibrium Loading (51 wt% CFC-113)
- SCFM to Adsorber (435 SCFM)
- Required Amount of Carbon (380 lb)
- Superficial Velocity (67 ft/min)
- Bed Area (6.8 ft^2)
- Bed Depth (1.8 ft)

TABLE D-2. EXAMPLE COST CALCULATIONS

A. Carbon Adsorber Only

- Given: Inlet Flow = 2500 ACFM (2196 scfm)

  Inlet Conc = 1000 ppmv CFC-113

1. Estimate Capital Cost of Adsorber Vessels: (From Ref. 10)

   $DCC_a = 155,600 + 1394\ A_t - 0.5996\ A_t^2$

   where: DCC_a = Direct capital cost of adsorber vessels ($).

   A_t = Total adsorber system bed area (31 ft^2)

   DCC_a = $152,554

2. Estimate Capital Cost of Duct Work: (From Ref. 10)

   $DCC_d = -2.35 \times 10^{-4}\ G^2 + 2.782G + 1205$ (for $G \leq 4000$)

   where: DCC_d = Direct capital cost of duct work ($)

   G = Vent stream flowrate at adsorber inlet (2196 scfm)

   DCC_d = $5,769

3. Estimate Capital Cost of Carbon:

   $DCC_c = d(A_t)(D)(U_{nc})$

   where: DCC_c = Direct capital cost of carbon ($)

   d = Bed depth (1.5 ft)

   A_t = Total adsorber system bed area (31 ft^2)

   D = Carbon density (30 lb/ft^3)

   U_{nc} = Unit cost of new carbon at $2.13/lb.

   DCC_c = $2,976

(Continued)

TABLE D-2. (Continued)

4. Estimate Capital Cost of Fans:

$P = (1.107 \times 10^{-4} V_s^2 + 0.03679\ V_s)d + 1.0$

where: P = Pressure drop across adsorber system, "H_2O

V_s = Superficial velocity (75 ft/min)

d = Bed depth (1.5 ft)

P = 6.07 "H_2O

$DCC_f = 5.711 \times 10^{-2}\ (P)(G) + 1965$

where: DCC_f = Direct capital cost of fans ($)

P = Pressure drop (6.07 "H_2O)

G = Vent stream flowrate (2196 scfm)

DCC_f = $2,727

5. Total Installed Direct Capital Cost:

$TEC = DCC_a + DCC_d + DCC_c + DCC_f$

where: TEC = Total installed equipment cost of adsorber system ($)
DCC_a = Cost of adsorber vessels ($152,554)
DCC_d = Cost of duct work ($5,769)
DCC_c = Cost of carbon ($2,976)
DCC_f = Cost of fans ($2,727)

TEC = $164,026

6. Annual Cost for Capital Recovery:

$AC_{cr} = (CRF)(TEC)$

where: AC_{cr} = Annual cost for capital recovery ($)

CRF = Capital recovery factor (for 10% interest over 10 years = 0.1627)

TEC = Total equipment cost ($164,026)

AC_{cr} = $26,687

(Continued)

TABLE D-2. (Continued)

7. Annual Cost for Carbon Replacement:

$AC_{cb} = DCC_c/L$

where: AC_{cb} = Annual carbon replacement cost ($)

DCC_c = Capital cost of carbon ($2,976)

L = Bed life (3 years)

AC_{cb} = $992

8. Annual Cost for Electricity:

$HP_f = 2.426 \times 10^{-4} (P)(G)$

where: HP_f = Fan horsepower

P = Pressure drop across system (6.07 "H_2O)

G = Vent stream flowrate (2196 scfm)

HP_f = 3.23 HP

$AC_e = 0.7457 (V_e)(HP_f)(HR)$

where: AC_e = Annual cost for electricity ($)

V_e = Unit cost for electricity ($0.0572/kwh)

HP_f = Fan horsepower (3.23 HP)

HR = Annual hours of adsorber operation (8200 hrs/yr)

AC_e = $1,133

9. Annual Cost for Maintenance:

$h_m = 77.5 + 1.25 \times 10^{-3} (G)$

where: h_m = Hours of maintenance labor

G = Vent stream flowrate (2196 scfm)

h_m = 80 hrs

(Continued)

TABLE D-2. (Continued)

$AC_m = 2.5\ (U_{ml} \times h_m)$

where: AC_m = Annual maintenance cost

U_{ml} = Maintenance labor cost ($19.07/hr)

h_m = Hours of maintenance labor (80 hours)

AC_m = $3,826

10. Annual Cost for Operating Labor:

$AC_{ol} = 365\ U_{ol}\ HR/8760$

where: AC_{ol} = Annual cost for operating labor ($)

U_{ol} = Unit cost for operating labor ($28.61/day)

HR = Annual hours of adsorber operation (8200 hrs/yr)

AC_{ol} = $9,785 (Same for all cases)

11. Annual Cost for Steam:

$S_r = (R_t)/2$

where: S_r = Steam required per regeneration (lb)

R_t = Total amount of carbon (1398 lb)

S_r = 699 lb/regeneration (Equivalent to 1 lb steam per lb of carbon in bed)

$AC_s = (S_r)(HR/t)(U_s)$

where: AC_s = Annual steam cost ($)

S_r = Steam required per regeneration (699 lb)

HR = Annual hours of adsorber operation (8200 hrs/yr)

t = Hours per regeneration (4.91 hours)

U_s = Unit cost of steam ($0.0055/lb)

AC_s = $6,425

(Continued)

TABLE D-2. (Continued)

12. Annual Credit for Recovered Solvent:

$SL_r = 1.556 \times 10^{-7} (MW)(C)(G)(.95)$

where: SL_r = Solvent recovery rate (lb/hr)
MW = Molecular weight of solvent (187.4 lb/mol)
C = Concentration (1000 ppmv)
G = Vent stream flowrate (2196 scfm)

SL_r = 60.8 lb/hr

$AC_p = (SL_r)(U_{sl}/D_s)(0.5)(HR)$

where: AC_p = Annual recovered solvent credit
SL_r = Solvent recovery rate (60.8 lb/hr)
U_{sl} = Unit cost of virgin solvent ($1/gal)
D_s = Density of virgin solvent (8.0 lb/gal)
HR = Annual hours of adsorber operation (8200 hrs/yr)

AC_p = <$33,526> (Assumes recovered solvent is worth 50% of virgin solvent)

13. Annual Cost for Cooling Water:

$AC_{cw} = 5.36 \times 10_{-3} (U_{cw})(S_r/t)(HR)$

where: AC_{cw} = Annual cost for cooling water
U_{cw} = Unit cost of cooling water
S_r = Steam required per regeneration (699 lb)
t = Time between regenerations (4.91 hr)
HR = Annual hours of adsorber operation (8200 hrs/yr)

AC_{cw} = $3,131

14. Annual Cost for Wastewater Disposal:

$AC_{ww} = 5.36 \times 10^{-3} (U_{ww})(S_r/t)(HR)$

where: AC_{ww} = Annual cost for disposal of wastewater
U_{ww} = Unit cost of disposal of wastewater
S_r = Steam required per regeneration (699 lb)
t = Time between regenerations (4.91 hr)
HR = Annual hours of operation (8200 hrs/yr)

AC_{ww} = $9,392

(Continued)

TABLE D-2. (Continued)

15. Total Annual Operating Costs:

$TOC = AC_{cb} + AC_{e} + AC_{m} + AC_{ol} + AC_{s} + AC_{cw} + AC_{ww} + AC_{p}$

(described above)

TOC = $1,158

16. Total Annual Costs:

$TAC = AC_{cr} + TOC$

where: TAC = Total annualized costs
AC_{cr} = Annual cost for capital recovery ($26,687)
TOC = Total operating costs ($1,158)

TAC = $27,866

17. Costs Per Metric Ton Controlled:

Metric Tons of Solvent Controlled = $(N_s)(CE_{ca})(HR)(MW)/2200$

where: N_s = Inlet solvent molar flow (0.367 lb mol/hr)
CE_{ca} = Control efficiency, carbon adsorption (95%)
HR = Annual operating hours (8200 hrs/yr)
MW = Solvent molecular weight (187.4 lb/lb mol)

Metric Tons Controlled = 244 MT/yr

Cost per Metric Ton = Total Annual Cost/Metric Tons Controlled

= $27,866/244 = $144/MT

B. Membrane Design with Carbon Adsorber (Using Vacuum Pump):

- Given: Membrane Selectivity (α^*)=20

  Pressure Ratio (8) = 0.10

  Removal Efficiency = 60%

(Continued)

TABLE D-2. (Continued)

18. Estimate Capital Cost of Vacuum Pump:

Installed Capital Cost = $34490 \sqrt{(HP/10)}$

where: HP = Required horsepower of vacuum pump (180 HP)

Installed Capital Cost = $146,125

19. Estimate Capital Cost of Membrane and Other Auxiliaries:

Membrane Capital Cost = $100(S)/(3.281)^2$

where: S = Required membrane area (27,400 ft^2)

Membrane Capital Cost = $259,848 (Based on $\$100/m^2$)

Auxiliary Equipment Capital Cost = $50(S)/(3.281)^2$

Auxiliary Equipment Capital Cost = $129,924 (Based on $\$50/m^2$)

20. Estimate Installed Capital Cost for Carbon Adsorber: (see steps 1-5 above)

DCC_a	= $124,703	(Based on required area of 4.7 ft)
DCC_d	= $4,546	(Based on flowrate of 278 scfm)
DCC_c	= $814	(Based on required carbon of 353 lb)
DCC_f	= $1,965	(Based on velocity of 63 ft/min)

TEC = $132,028

21. Estimate Annual Operating Costs: (see steps 7-1 above)

AC_e = $63,031	(Based on vacuum pump HP = 180 and adsorber fan = 0.5)
AC_{cb} = $251	(Based on required carbon of 353 lb and three-year life)
AC_m = $3,711	(Based on 77.8 hrs of maintenance labor)
AC_{ol} = $9,785	(Same for all cases)
AC_s = $3,191	(Based on 177 lb/regen and 2.5 hrs between regenerations)

(Continued)

TABLE D-2. (Continued)

AC_{mem} = $86,616 (Based on three-year life)
AC_{cw} = $1,555
AC_{ww} = $4,665

TOC = $152,861

22. Estimate Annual Cost for Capital Recovery: (see step 6 above)

AC_{cr} = (CRF)(TEC)

where: CRF = 0.1627
TEC = $132,028 (adsorber)
$146,125 (vacuum pump)
$259,848 (membrane)
$129,924 (mem. auxil.)

TEC = $667,925

AC_{cr} = $108,671

23. Total Annual Cost:

TAC = AC_{cr} + TOC

= $261,533

24. Cost Per Metric Ton Controlled:

Metric Tons Controlled = $(N_s)(CE_{mem})(CE_{ca})(HR)(MW)/2200$

where: N_s = Inlet solvent molar flow (0.367 lb mol/hr)
CE_{mem} = Control efficiency, membrane (60%)
CE_{ca} = Control efficiency, carbon adsorber (95%)
HR = Annual operating hours (8200 hrs/yr)
MW = Solvent molecular weight (187.4 lb/lb mol)

Metric Tons Controlled = 146 MT/yr

Cost Per Metric Ton = Total Annual Cost/Metric Tons Controlled

= $261,533/146 = $1,803/MT

APPENDIX E

QUALITY CONTROL EVALUATION REPORT (QCER)

QUALITY CONTROL EVALUATION REPORT (QCER)

1. Overall Summary of Data Quality

The overall data goals of this project were achieved. The precision of the test data was high, with virtually all of the results falling within 95% tolerance limits. Accuracy could not be quantitatively assessed due to lack of a performance audit, but critical THC measurements appear consistent and reliable. Data completeness was good. Raw data was recorded in a bound laboratory notebook at five-minute intervals. Data representativeness was good. Care was taken in the design and sampling to ensure consistent performance of the bench-scale system. Data comparability was good. Test results from this study were checked by the membrane manufacturer and agreed with expectations.

The reliability of the data is good. For example, a replicate was attempted for methylene chloride with an inlet concentration of about 90 ppm. The original run of 1/5/89 achieved a removal efficiency of 60% and an enrichment ratio of 3.35, while the replicate run on 1/6/89 achieved a removal efficiency of 64.5% and an enrichment ratio of 2.92. The percent differences for removal efficiency and enrichment ratio were 7% and 13%, respectively. Other trends were consistent with expectations. For example, tests conducted with (greater vacuum) demonstrated higher enrichment ratios and removal efficiency.

It was determined that the gravimetric trap method did not meet the data quality objective (DQO) of greater than 95% recovery and, therefore, an alternate method was developed to close the material balance. The direct permeate measurement method showed much higher recovery and material balance closures.

2. Project QA Activities

The primary QA activity for this project was the preparation of a Test Plan/QAPP (EPA contract 68-02-4286, Work Assignment No. 32, Revised November 8, 1988). The initial QAPP was reviewed by Research Triangle Institute (for Judith Ford, EPA-AEERL) and was revised based on comments received.

A Systems Audit was performed on 1/19/89 which reviewed laboratory procedures for consistency with the QAPP. The auditor noted no problems with facilities and equipment, but found that a control standard was not purchased as planned. Other deficiencies concerned logbook recordkeeping, such as lack of initials to indicate supervisor review of calculation, and lack of explanation to corrections in logbook.

Undocumented changes to the QAPP included lack of drift checks, because a complete standard gas multipoint calibration was run daily, and failure to calculate the linear regression immediately after performing the THC analyzer calibration.

3. Detailed Discussion of Data Quality Indicators

Precision

The laboratory test results were subjected to a statistical analysis to assess the precision of the data. The statistical analysis consisted of the following calculations:

- Calculate average values of inlet solvent mass flow (in g/hr), outlet solvent mass flow, and to the extent possible, permeate solvent mass flow.

- Calculate the standard deviation of each data set using the equation below:

$$S = \Sigma (x_i - \bar{x})/n-1$$

where, C.V. = coefficient of variation;

x_i = i^{th} value of x;

$\bar{x}$ = Mean (average) value of x; and

h = number of x values.

- Calculate a 95% tolerance interval for each data set using the Factors for Two-sided Tolerance Limits for Normal Distributions (taken from Table E-1) and the equation shown below:

$$\bar{x} \pm K(s)$$

where, $\bar{x}$ = mean value of x;

k = factor for two-sided tolerance limit for normal distribution (95% probability that 95% of the distribution will be included); and

s = standard deviation of sample.

- Compare actual data values with 95% tolerance intervals. Data is acceptable if it falls with the tolerance interval. Table E-2 shows that only (2) of the (45) data sets had less than 100% of the data contained within the tolerance interval. We conclude that the overall precision of the data is excellent.

Accuracy/Bias

Difficult to quantify overall since a performance audit was not attempted. Qualitative estimates would be that accuracy was good because material balance closures were high (approaching 100%). Accuracy of some individual measurements can be estimated. Calibration data for the saturator thermocouple shows the indicated temperature to be approximately 3°F low of the true value. An appropriate correction was employed during data reduction. All other direct measurements (temperature, pressure, flow) were uncorrected.

TABLE E-1. FACTORS FOR TWO-SIDED TOLERANCE LIMITS FOR NORMAL DISTRIBUTIONS

Factors K such that the probability is γ that at least a proportion P of the distribution will be included between $\bar{x} \pm Ks$, where $\bar{x}$ and s are estimates of the mean and the standard deviation computed from a sample size of n.

$\gamma = 0.95$
P = 0.95

n	K
2	37.674
3	9.916
4	6.370
5	5.079
6	4.414
7	4.007
8	3.732
9	3.532
10	3.379
11	3.259
12	3.162
13	3.081
14	3.012
15	2.954
16	2.903
17	2.858
18	2.819
19	2.784
20	2.752
21	2.723
22	2.697
23	2.673
24	2.651
25	2.631
26	2.612
27	2.595

TABLE E-2. NINETY-FIVE PERCENT TOLERANCE INTERVALS AND DATA QUALITY

Solvent	Date	Inlet Mass Flow (g/hr) 95% Tolerance Interval	Inlet % of Data Within Interval	Outlet Mass Flow (g/hr) 95% Tolerance Interval	Outlet % of Data Within Interval	Permeate Mass Flow (g/hr) 95% Tolerance Interval	Permeate % of Data Within Interval
Hexane	12-14-88	1.493 < x < 1.806	100%	0.4575 < x < 0.5385	100%		
	12-14-88	7.007 < x < 8.472	100%	1.9460 < x < 2.6760	100%		
	1-20-89	0.3076 < x < 0.3504	100%	0.1230 < x < 0.1410	100%	0.1565 < x < 0.1835	100%
	1-20-89	3.442 < x < 3.886	100%	1.431 < x < 1.639	100%	1.7073 < x < 2.0750	100%
Toluene	1-19-89	0.1708 < x < 0.1872	100%	0.0670 < x < 0.0718	100%	0.0790 < x < 0.1120	100%
	1-19-89	0.6402 < x < 0.7418	100%	0.2155 < x < 0.2885	100%	0.3295 < x < 0.4465	100%
MEK	12-16-88	0.1874 < x < 0.3366	100%	0.0368 < x < 0.0892	100%		
	12-19-88	1.318 < x < 1.624	100%	0.4660 < x < 0.7040	95%		
	1-3-89	4.779 < x < 10.089	100%	1.402 < x < 4.080	100%		
MeOH	1-13-89	0.1575 < x < 0.1625	100%	0.0811 < x < 0.0883	92%	0.0861 < x < 0.1092	100%
	1-13-89	0.6975 < x < 0.7205	100%	0.2841 < x < 0.3165	100%	0.2663 < x < 0.5393	100%
	1-16-89	1.949 < x < 2.513	100%	0.7817 < x < 1.0103	100%	1.113 < x < 1.593	100%
CFC-113	1-9-89	0.1550 < x < 0.2164	100%	0.0696 < x < 0.1364	100%		
	1-9-89	1.455 < x < 1.825	100%	0.6853 < x < 0.8357	100%		
	1-10-89	14.044 < x < 16.364	100%	5.357 < x < 7.181	100%		
MeCl	1-5-89	0.3504 < x < 0.5576	100%	0.1347 < x < 0.2289	100%		
	1-6-89	0.3526 < x < 0.5216	100%	0.1351 < x < 0.1747	100%	0.0963 < x < 0.3791	100%
	1-6-89	5.634 < x < 6.836	100%	2.186 < x < 2.508	100%	3.879 < x < 3.937	100%

Completeness

Data capture was very good as essentially all of the test data were usable. There were no instrumentation failures which could have invalidated any of the runs.

Data completeness could have been improved even further with the use of a computer-based data acquisition system. However, this was not available for the current study.

Representativeness

Data representativeness was good. The design of the test system and sampling procedures took account of possible problems which might limit the utility of data from the project. For instance, the use of nitrogen as a carrier gas in the system prevented possible flammability concerns and allowed better comparison with work by others in the field. Materials of construction were selected for inertness, so Teflon and stainless steel were used. Blanks were run before each new solvent test series to ensure that all traces of solvent from previous tests had been completely purged.

Comparability

Review of preliminary test results by the membrane manufacturer confirmed that the test data was meeting expectations.

Comparison with other laboratory test data was more difficult, since the only other data using the particular membrane unit was data supplied by the manufacturer. The manufacturer's data were obtained at higher concentrations (i.e., volume percent levels instead of ppm). Nevertheless, the comparison was important in that the correct relationship between both data sets was observed.

Figures E-1 through E-3 illustrate the relationship between the manufacturer's data (boxes) versus the present work (crosses) for toluene, MEK, and MeCl, respectively. The expected trend would be for the line to pass through the origin, since at zero concentration, the flux should be zero. The comparability appears to be excellent.

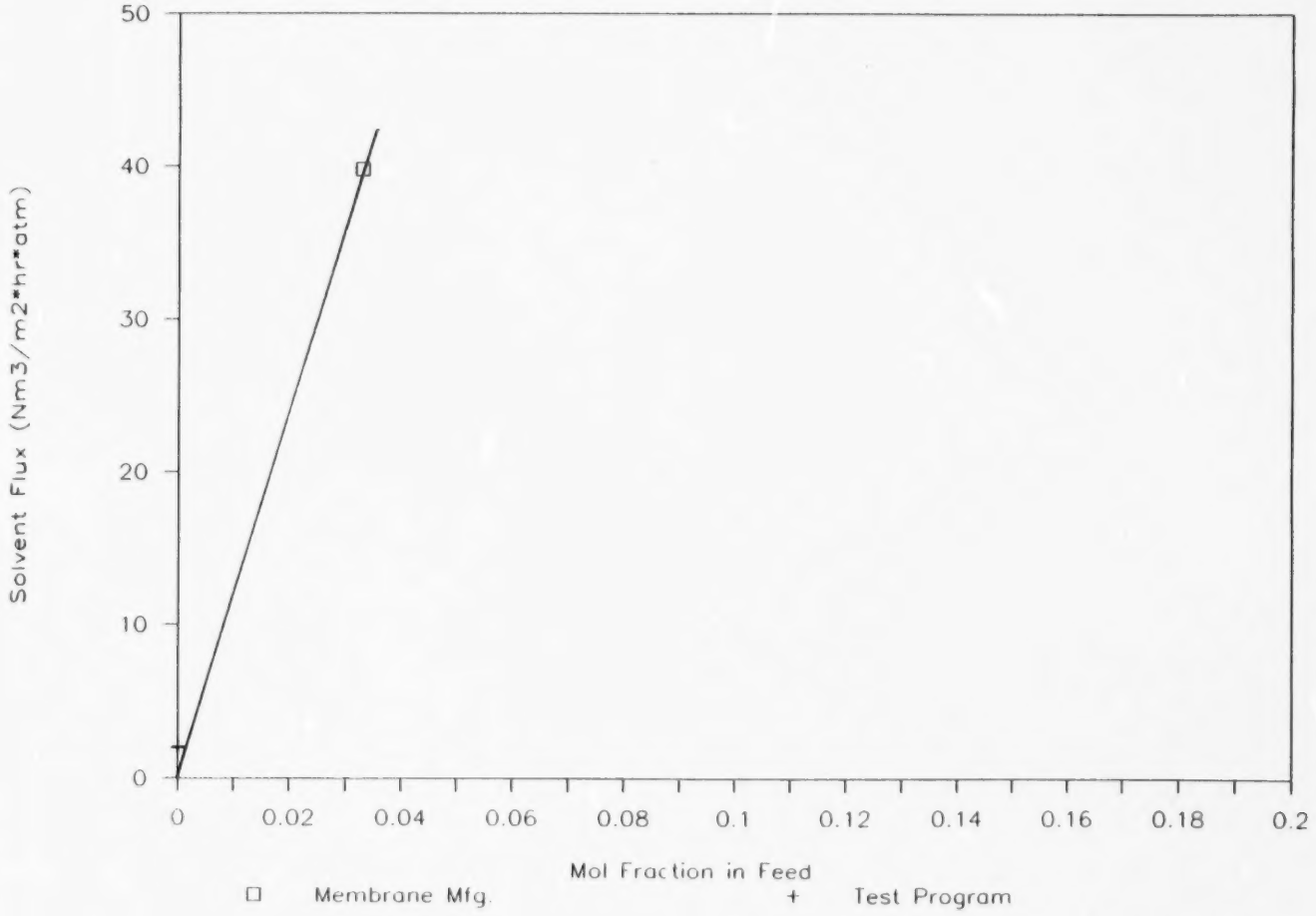

Figure E-1. Comparison of Manufacturer's Data to Test Data (Toluene)

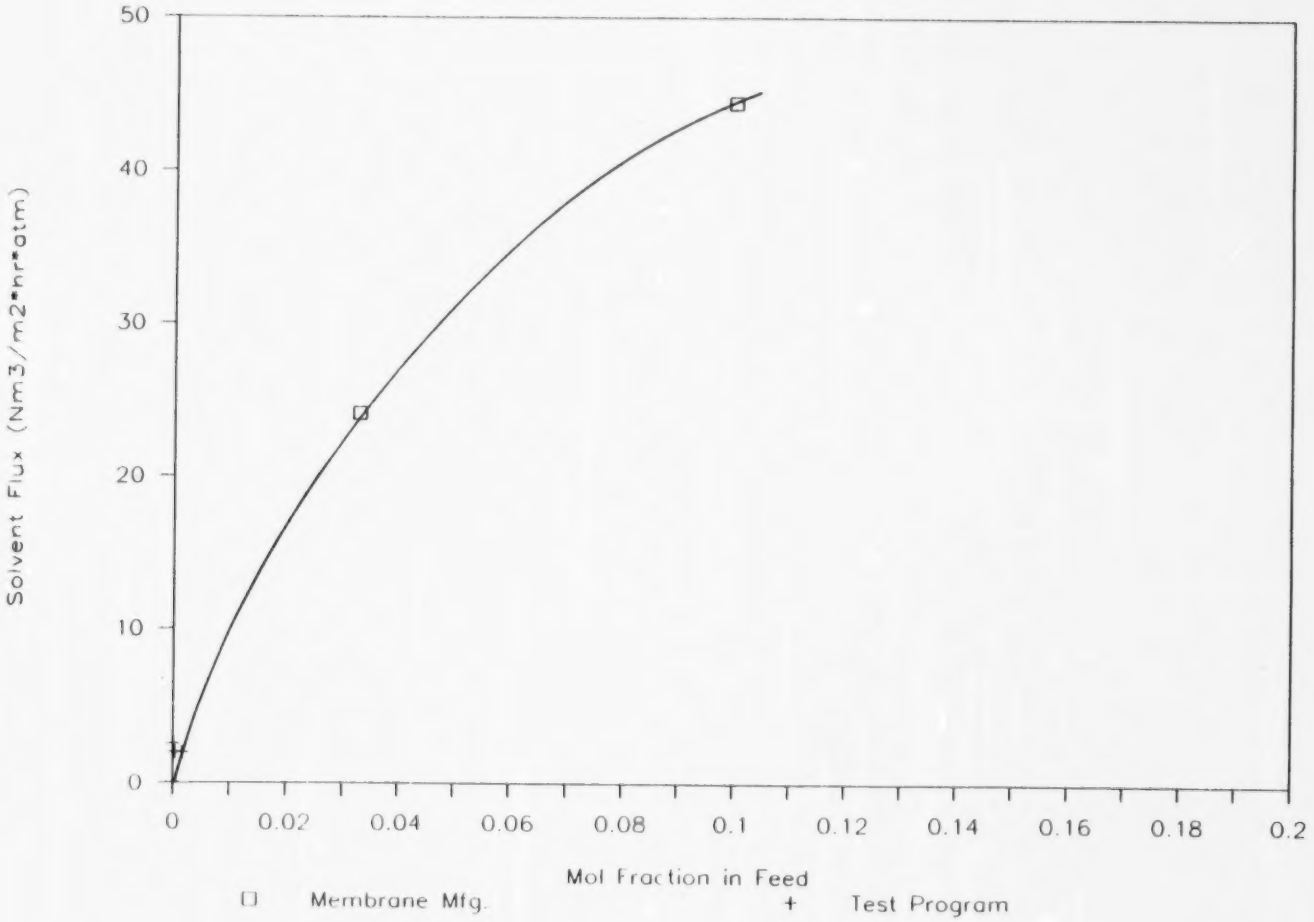

Figure E-2. Comparison of Manufacturer's Data to Test Data (Methyl Ethyl Ketone)

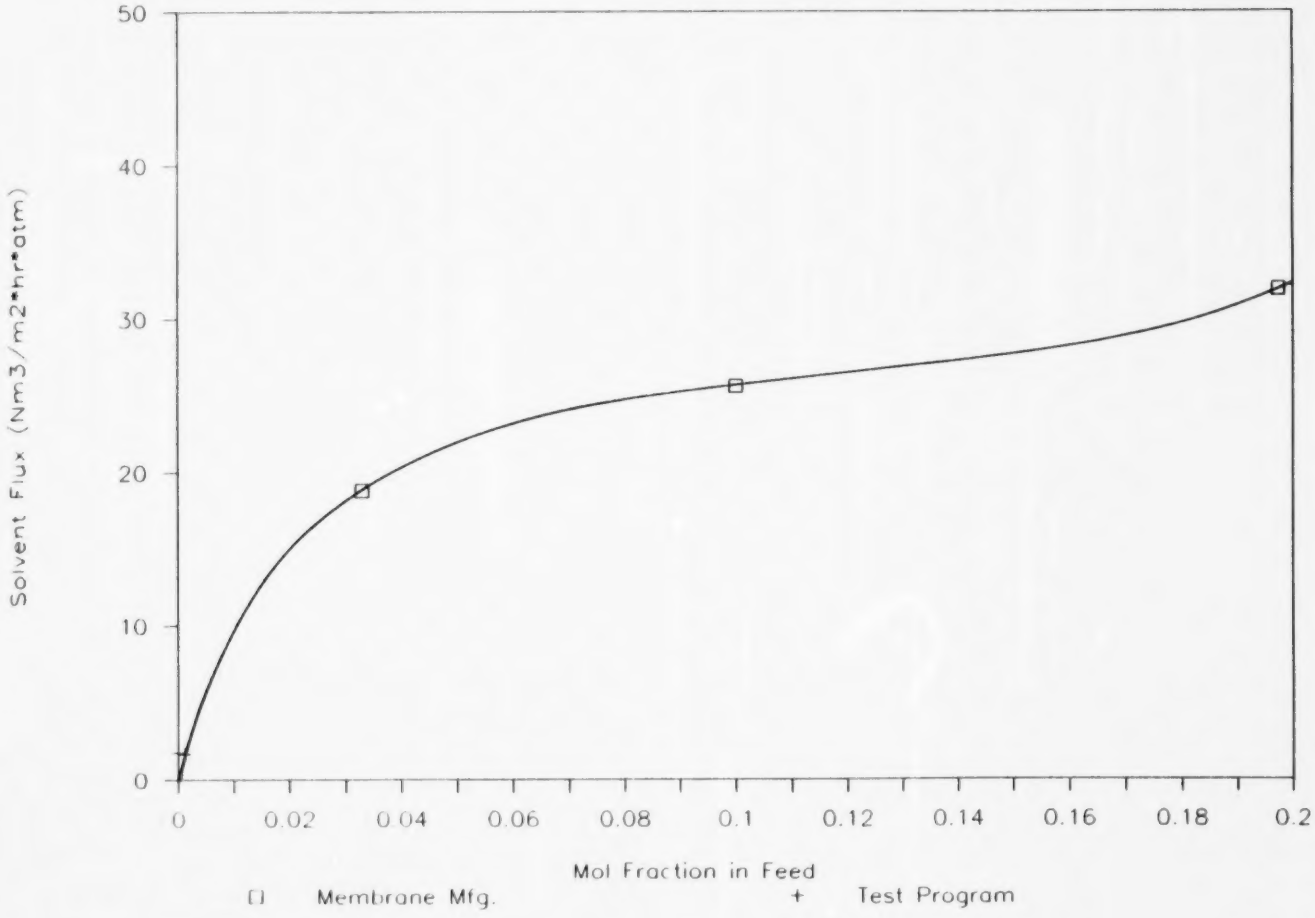

Figure E-3. Comparison of Manufacturer's Data to Test Data (Methylene Chloride)

APPENDIX F
DETAILED DESIGN CALCULATIONS

TABLE F-1. SPREADSHEET CALCULATIONS FOR SYSTEM DESIGN

INPUT DATA

CASE No.	Inlet Flow (acfm)	Solvent Inlet Conc. (ppmv)	Solvent Name	Solvent Molec. Wt.	Inlet Temp. (deg F)	Inlet RH (%)	Total Inlet Molar Flow (lbmol/min)	Total Solvent Inlet Molar Flow (lbmol/min)	(lbmol/hr)
1	250	1000	CFC-113	187.4	100	18	0.6118	0.00061	0.037
2	2500	1000	CFC-113	187.4	100	18	6.1182	0.00612	0.367
3	10000	1000	Toluene	92.1	100	18	24.4727	0.02447	1.468
4	250	100	CFC-113	187.4	100	18	0.6118	0.00006	0.004
5	2500	100	CFC-113	187.4	100	18	6.1182	0.00061	0.037
6	10000	100	Toluene	92.1	100	18	24.4727	0.00245	0.147

A. BASE CASE: Direct Treatment using Carbon Adsorption

CARBON ADSORPTION

	FLOW (SCFM)	CONC (PPMV)	Reqd Amt of Carbon (pounds)	Superficial Velocity (ft/min)	Bed Area Normal to Flow (ft^2)	Bed Depth (ft)	new Rt (@ 1.5ft)	new t	Total Area (ft^2)
CASE 1	219.6	1000	71	75	3	0.764	139.8	4.91	3
CASE 2	2196.4	1000	712	75	31	0.764	1397.7	4.91	31
CASE 3	8785.7	1000	2098	75	124	0.563	5590.9	6.66	186
CASE 4	219.6	100	18	100	2	0.255	104.8	14.73	2
CASE 5	2196.4	100	178	100	23	0.255	1048.3	14.73	23
CASE 6	8785.7	100	262	100	93	0.094	4193.2	39.96	140

TABLE F-1. (Continued)

B. Membrane Preconcentrator with Vacuum Pump and Carbon Adsorber

	FLOW (SCFM)	CONC (PPMV)
CASE 1	219.6	1000

Approx. Weller & Steiner Solution (xf-->0)

Constants: ========= Calculated: =========

alpha*	gamma	xf	x	F	y	Rf	xw/xf	% Removal	Solvent Flux (lbmol/hr)	Area (ft^2)	Gas Flow to Adsorb. (lbmol/min)	ACFM to Vac. Pump	Required HP for Vac. Pump	Enrichmt. Ratio
200	0.1	0.001	0.00044	0.091357	0.00656973	0.101508	0.44	60.02	0.0245	20178	0.056	228	13	6.570
200	0.1	0.001	0.000175	0.184044	0.00465761	0.204493	0.175	85.72	0.0350	40649	0.113	460	26	4.658
200	0.1	0.001	0.00006	0.279860	0.00341882	0.310955	0.06	95.68	0.0390	61811	0.171	700	40	3.419
20	0.1	0.001	0.00045	0.126650	0.00479266	0.140722	0.45	60.70	0.0248	2797	0.077	317	18	4.793
20	0.1	0.001	0.000195	0.242126	0.00351970	0.269029	0.195	85.22	0.0348	5348	0.148	605	34	3.520
20	0.1	0.001	0.00007	0.362999	0.00263198	0.403333	0.07	95.54	0.0390	8017	0.222	907	51	2.632
5	0.1	0.001	0.000515	0.227452	0.00264731	0.252725	0.515	60.21	0.0246	1256	0.139	569	32	2.647
5	0.1	0.001	0.00025	0.416735	0.00204970	0.463039	0.25	85.42	0.0348	2301	0.255	1042	59	2.050
5	0.1	0.001	0.00011	0.576153	0.00165472	0.640170	0.11	95.34	0.0389	3181	0.353	1440	82	1.655

	FLOW (SCFM)	CONC (PPMV)
CASE 2	2196.4	1000

Approx. Weller & Steiner Solution (xf-->0)

Constants: ========= Calculated: =========

alpha*	gamma	xf	x	F	y	Rf	xw/xf	% Removal	Solvent Flux (lbmol/hr)	Area (ft^2)	Gas Flow to Adsorb. (lbmol/min)	ACFM to Vac. Pump	Required HP for Vac. Pump	Enrichmt. Ratio
200	0.1	0.001	0.00044	0.091357	0.00656973	0.101508	0.44	60.02	0.2448	201777	0.559	2284	129	6.570
200	0.1	0.001	0.000175	0.184044	0.00465761	0.204493	0.175	85.72	0.3496	406488	1.126	4601	261	4.658
200	0.1	0.001	0.00006	0.279860	0.00341882	0.310955	0.06	95.68	0.3903	618111	1.712	6997	397	3.419
20	0.1	0.001	0.00045	0.126650	0.00479266	0.140722	0.45	60.70	0.2476	27973	0.775	3166	180	4.793
20	0.1	0.001	0.000195	0.242126	0.00351970	0.269029	0.195	85.22	0.3476	53477	1.481	6053	343	3.520
20	0.1	0.001	0.00007	0.362999	0.00263198	0.403333	0.07	95.54	0.3897	80174	2.221	9075	514	2.632
5	0.1	0.001	0.000515	0.227452	0.00264731	0.252725	0.515	60.21	0.2456	12559	1.392	5686	322	2.647
5	0.1	0.001	0.00025	0.416735	0.00204970	0.463039	0.25	85.42	0.3484	23010	2.550	10418	591	2.050
5	0.1	0.001	0.00011	0.576153	0.00165472	0.640170	0.11	95.34	0.3889	31813	3.525	14404	817	1.655

TABLE F-1. (Continued)

B. Membrane Preconcentrator with Vacuum Pump and Carbon Adsorber

	FLOW (SCFM)	CONC (PPMV)
CASE 1	219.6	1000

alpha*	% Removal	SCFM to Carbon Adsorber	Reqd Amt of Carbon (pounds)	Superficial Velocity (ft/min)	Bed Area Normal to Flow (ft^2)	Bed Depth (ft)	new Ao (@ 3.0ft)	new Vs	Total Area (ft^2)
200	60.02	20.07	34	59	0.4	3.149	0.4	55.81	0.4
200	85.72	40.42	50	63	0.7	2.458	0.6	77.31	0.7
200	95.68	61.47	59	66	1.0	2.002	0.7	99.58	1.0
20	60.70	27.82	35	63	0.5	2.516	0.4	75.13	0.5
20	85.22	53.18	52	66	0.9	2.053	0.6	96.73	0.9
20	95.54	79.73	61	68	1.2	1.650	0.7	124.38	1.2
5	60.21	49.96	39	68	0.8	1.659	0.4	123.66	0.8
5	85.42	91.53	57	70	1.4	1.367	0.6	153.32	1.4
5	95.34	126.55	65	71	1.9	1.143	0.7	185.96	1.9

	FLOW (SCFM)	CONC (PPMV)
CASE 2	2196.4	1000

alpha*	% Removal	SCFM to Carbon Adsorber	Reqd Amt of Carbon (pounds)	Superficial Velocity (ft/min)	Bed Area Normal to Flow (ft^2)	Bed Depth (ft)	new Ao (@ 3.0ft)	new Vs	Total Area (ft^2)
200	60.02	200.66	343	59	3.6	3.149	3.8	55.81	3.6
200	85.72	404.24	499	63	6.8	2.458	5.5	77.31	6.8
200	95.68	614.69	589	66	9.8	2.002	6.5	99.58	9.8
20	60.70	278.18	353	63	4.7	2.516	3.9	75.13	4.7
20	85.22	531.81	525	66	8.5	2.053	5.8	96.73	8.5
20	95.54	797.30	612	68	12.4	1.650	6.8	124.38	12.4
5	60.21	499.58	386	68	7.7	1.659	4.3	123.66	7.7
5	85.42	915.33	570	70	13.9	1.367	6.3	153.32	13.9
5	95.34	1265.48	650	71	18.9	1.143	7.2	185.96	18.9

TABLE F-1. (Continued)

CASE 3 8785.7 1000

Approx. Weller & Steiner Solution (xf-->0)

Constants				Calculated					Solvent Flux		Gas Flow to Adsorb.	ACFM	Required HP	Enrichmt.
alpha*	gamma	xf	x	F	y	Rf	xw/xf	% Removal	(lbmol/hr)	Area (ft^2)	(lbmol/min)	to Vac. Pump	for Vac. Pump	Ratio
200	0.1	0.001	0.00044	0.091357	0.00656973	0.101508	0.44	60.02	0.9792	807107	2.236	9136	518	6.570
200	0.1	0.001	0.000175	0.184044	0.00465761	0.204493	0.175	85.72	1.3985	1625950	4.504	18404	1043	4.658
200	0.1	0.001	0.00006	0.279860	0.00341882	0.310955	0.06	95.68	1.5610	2472442	6.849	27986	1587	3.419
20	0.1	0.001	0.00045	0.126650	0.00479266	0.140722	0.45	60.70	0.9903	111890	3.099	12665	718	4.793
20	0.1	0.001	0.000195	0.242126	0.00351970	0.269029	0.195	85.22	1.3904	213909	5.926	24213	1373	3.520
20	0.1	0.001	0.00007	0.362999	0.00263198	0.403333	0.07	95.54	1.5588	320695	8.884	36300	2058	2.632
5	0.1	0.001	0.000515	0.227452	0.00264731	0.252725	0.515	60.21	0.9824	50236	5.566	22745	1289	2.647
5	0.1	0.001	0.00025	0.416735	0.00204970	0.463039	0.25	85.42	1.3936	92042	10.199	41674	2363	2.050
5	0.1	0.001	0.00011	0.576153	0.00165472	0.640170	0.11	95.34	1.5554	127252	14.100	57615	3266	1.655

CASE 4 219.6 100

Approx. Weller & Steiner Solution (xf-->0)

Constants				Calculated					Solvent Flux		Gas Flow to Adsorb.	ACFM	Required HP	Enrichmt.
alpha*	gamma	xf	x	F	y	Rf	xw/xf	% Removal	(lbmol/hr)	Area (ft^2)	(lbmol/min)	to Vac. Pump	for Vac. Pump	Ratio
200	0.1	0.0001	0.000044	0.091357	0.00065697	0.101508	0.44	60.02	0.0024	20178	0.056	228	13	6.570
200	0.1	0.0001	0.000017	0.184044	0.00046576	0.204493	0.175	85.72	0.0035	40649	0.113	460	26	4.658
200	0.1	0.0001	0.000006	0.279860	0.00034188	0.310955	0.06	95.68	0.0039	61811	0.171	700	40	3.419
20	0.1	0.0001	0.000045	0.126650	0.00047926	0.140722	0.45	60.70	0.0025	2797	0.077	317	18	4.793
20	0.1	0.0001	0.000019	0.242126	0.00035197	0.269029	0.195	85.22	0.0035	5348	0.148	605	34	3.520
20	0.1	0.0001	0.000007	0.362999	0.00026319	0.403333	0.07	95.54	0.0039	8017	0.222	907	51	2.632
5	0.1	0.0001	0.000051	0.227452	0.00026473	0.252725	0.515	60.21	0.0025	1256	0.139	569	32	2.647
5	0.1	0.0001	0.000025	0.416735	0.00020497	0.463039	0.25	85.42	0.0035	2301	0.255	1042	59	2.050
5	0.1	0.0001	0.000011	0.576153	0.00016547	0.640170	0.11	95.34	0.0039	3181	0.353	1440	82	1.655

TABLE F-1. (Continued)

CASE 3 8785.7 1000

alpha*	% Removal	SCFM to Carbon Adsorber	Reqd Amt of Carbon (pounds)	Superficial Velocity (ft/min)	Bed Area Normal to Flow (ft^2)	Bed Depth (ft)	new Ao (@ 3.0ft)	new Vs	Total Area (ft^2)
200	60.02	802.64	1021	59	14.5	2.342	11.3	75.03	14.5
200	85.72	1616.96	1542	63	27.1	1.899	17.1	100.11	27.1
200	95.68	2458.77	1772	66	39.2	1.505	19.7	132.48	39.2
20	60.70	1112.71	1092	63	18.7	1.943	12.1	97.29	18.7
20	85.22	2127.26	1578	66	34.1	1.543	17.5	128.69	34.1
20	95.54	3189.21	1823	68	49.4	1.229	20.3	167.03	49.4
5	60.21	1998.34	1149	68	31.0	1.235	12.8	166.06	31.0
5	85.42	3661.32	1680	70	55.6	1.008	18.7	207.98	55.6
5	95.34	5061.92	1936	71	75.8	0.852	21.5	249.57	113.6

CASE 4 219.6 100

alpha*	% Removal	SCFM to Carbon Adsorber	Reqd Amt of Carbon (pounds)	Superficial Velocity (ft/min)	Bed Area Normal to Flow (ft^2)	Bed Depth (ft)	new Rt (@ 1.5ft)	new t	Total Area (ft^2)
200	60.02	20.07	5	91	0.2	0.654	10.5	5.74	0.2
200	85.72	40.42	7	94	0.5	0.503	20.5	7.46	0.5
200	95.68	61.47	8	96	0.7	0.397	30.5	9.46	0.7
20	60.70	27.82	5	94	0.3	0.516	14.1	7.27	0.3
20	85.22	53.18	7	96	0.6	0.408	26.4	9.20	0.6
20	95.54	79.73	8	97	0.9	0.318	39.1	11.81	0.9
5	60.21	49.96	5	97	0.5	0.319	24.5	11.74	0.5
5	85.42	91.53	8	98	1.0	0.264	44.4	14.21	1.0
5	95.34	126.55	9	99	1.4	0.221	61.0	16.99	1.4

TABLE F-1. (Continued)

CASE 5 2196.4 100

Approx. Weller & Steiner Solution (xf-->0)

Constants

alpha*	gamma	xf	x	Calculated F	y	Rf	xw/xf	% Removal	Solvent Flux (lbmol/hr)	Area (ft^2)	Gas Flow to Adsorb. (lbmol/min)	ACFM to Vac. Pump	Required HP for Vac. Pump	Enrichmt. Ratio
200	0.1	0.0001	0.000044	0.091357	0.00065697	0.101508	0.44	60.02	0.0245	201777	0.559	2284	129	6.570
200	0.1	0.0001	0.000017	0.184044	0.00046576	0.204493	0.175	85.72	0.0350	406488	1.126	4601	261	4.658
200	0.1	0.0001	0.000006	0.279860	0.00034188	0.310955	0.06	95.68	0.0390	618111	1.712	6997	397	3.419
20	0.1	0.0001	0.000045	0.126650	0.00047926	0.140722	0.45	60.70	0.0248	27973	0.775	3166	180	4.793
20	0.1	0.0001	0.000019	0.242126	0.00035197	0.269029	0.195	85.22	0.0348	53477	1.481	6053	343	3.520
20	0.1	0.0001	0.000007	0.362999	0.00026319	0.403333	0.07	95.54	0.0390	80174	2.221	9075	514	2.632
5	0.1	0.0001	0.000051	0.227452	0.00026473	0.252725	0.515	60.21	0.0246	12559	1.392	5686	322	2.647
5	0.1	0.0001	0.000025	0.416735	0.00020497	0.463039	0.25	85.42	0.0348	23010	2.550	10418	591	2.050
5	0.1	0.0001	0.000011	0.576153	0.00016547	0.640170	0.11	95.34	0.0389	31813	3.525	14404	817	1.655

CASE 6 8785.7 100

Approx. Weller & Steiner Solution (xf-->0)

Constants

alpha*	gamma	xf	x	Calculated F	y	Rf	xw/xf	% Removal	Solvent Flux (lbmol/hr)	Area (ft^2)	Gas Flow to Adsorb. (lbmol/min)	ACFM to Vac. Pump	Required HP for Vac. Pump	Enrichmt. Ratio
200	0.1	0.0001	0.000044	0.091357	0.00065697	0.101508	0.44	60.02	0.0979	807107	2.236	9136	518	6.570
200	0.1	0.0001	0.000017	0.184044	0.00046576	0.204493	0.175	85.72	0.1399	1625950	4.504	18404	1043	4.658
200	0.1	0.0001	0.000006	0.279860	0.00034188	0.310955	0.06	95.68	0.1561	2472442	6.849	27986	1587	3.419
20	0.1	0.0001	0.000045	0.126650	0.00047926	0.140722	0.45	60.70	0.0990	111890	3.099	12665	718	4.793
20	0.1	0.0001	0.000019	0.242126	0.00035197	0.269029	0.195	85.22	0.1390	213909	5.926	24213	1373	3.520
20	0.1	0.0001	0.000007	0.362999	0.00026319	0.403333	0.07	95.54	0.1559	320695	8.884	36300	2058	2.632
5	0.1	0.0001	0.000051	0.227452	0.00026473	0.252725	0.515	60.21	0.0982	50236	5.566	22745	1289	2.647
5	0.1	0.0001	0.000025	0.416735	0.00020497	0.463039	0.25	85.42	0.1394	92042	10.199	41674	2363	2.050
5	0.1	0.0001	0.000011	0.576153	0.00016547	0.640170	0.11	95.34	0.1555	127252	14.100	57615	3266	1.655

TABLE F-1. (Continued)

CASE 5 2196.4 100

alpha*	% Removal	SCFM to Carbon Adsorber	Reqd Amt of Carbon (pounds)	Superficial Velocity (ft/min)	Bed Area Normal to Flow (ft^2)	Bed Depth (ft)	new Rt (@ 1.5ft)	new t	Total Area (ft^2)
200	60.02	200.66	46	91	2.3	0.654	105.0	5.74	2.3
200	85.72	404.24	69	94	4.6	0.503	204.8	7.46	4.6
200	95.68	614.69	81	96	6.8	0.397	305.0	9.46	6.8
20	60.70	278.18	49	94	3.1	0.516	141.2	7.27	3.1
20	85.22	531.81	72	96	5.9	0.408	264.3	9.20	5.9
20	95.54	797.30	83	97	8.7	0.318	390.6	11.81	8.7
5	60.21	499.58	52	97	5.4	0.319	244.8	11.74	5.4
5	85.42	915.33	78	98	9.9	0.264	444.2	14.21	9.9
5	95.34	1265.48	90	99	13.6	0.221	610.3	16.99	13.6

CASE 6 8785.7 100

alpha*	% Removal	SCFM to Carbon Adsorber	Reqd Amt of Carbon (pounds)	Superficial Velocity (ft/min)	Bed Area Normal to Flow (ft^2)	Bed Depth (ft)	new Rt (@ 1.5ft)	new t	Total Area (ft^2)
200	60.02	802.64	130	91	9.3	0.465	420.0	8.06	9.3
200	85.72	1616.96	193	94	18.2	0.353	819.0	10.62	18.2
200	95.68	2458.77	223	96	27.1	0.274	1220.1	13.67	27.1
20	60.70	1112.71	136	94	12.6	0.362	564.9	10.35	12.6
20	85.22	2127.26	199	96	23.5	0.282	1057.3	13.30	23.5
20	95.54	3189.21	223	97	34.7	0.214	1562.4	17.53	34.7
5	60.21	1998.34	140	97	21.8	0.215	979.2	17.44	21.8
5	85.42	3661.32	207	98	39.5	0.175	1776.9	21.48	39.5
5	95.34	5061.92	240	99	54.2	0.148	2441.2	25.42	81.4

TABLE F-1. (Continued)

C. Membrane Preconcentrator with Compressor and Carbon Adsorber

	FLOW (SCFM)	CONC (PPMV)
CASE 1	219.6	1000

Approx. Weller & Steiner Solution (xf-->0)

Constants				Calculated					Solvent Flux		Gas Flow to Adsorb.	ACFM	Required HP	Enrichmt.
alpha*	gamma	xf	x	F	y	Rf	xw/xf	% Removal	(lbmol/hr)	Area (ft^2)	(lbmol/min)	to Compressor	for Compr.	Ratio
200	0.2	0.001	0.000481	0.171028	0.00351558	0.213785	0.481	60.13	0.0276	8327	0.105	250	33	3.516
200	0.2	0.001	0.000217	0.323999	0.00263366	0.404999	0.217	85.33	0.0392	15775	0.198	250	33	2.634
200	0.2	0.001	0.000086	0.466748	0.00204422	0.583436	0.086	95.41	0.0438	22725	0.286	250	33	2.044
20	0.2	0.001	0.000497	0.198114	0.00303594	0.247642	0.497	60.15	0.0276	965	0.121	250	33	3.036
20	0.2	0.001	0.000235	0.367020	0.00231935	0.458775	0.235	85.12	0.0391	1787	0.225	250	33	2.319
20	0.2	0.001	0.000096	0.522897	0.00182482	0.653621	0.096	95.42	0.0438	2546	0.320	250	33	1.825
5	0.2	0.001	0.000555	0.281933	0.00213338	0.352416	0.555	60.15	0.0276	343	0.172	250	33	2.133
5	0.2	0.001	0.00029	0.501576	0.00170553	0.626970	0.29	85.55	0.0393	611	0.307	250	33	1.706
5	0.2	0.001	0.00014	0.669099	0.00142530	0.836374	0.14	95.37	0.0438	814	0.409	250	33	1.425

	FLOW (SCFM)	CONC (PPMV)
CASE 2	2196.4	1000

Approx. Weller & Steiner Solution (xf-->0)

Constants				Calculated					Solvent Flux		Gas Flow to Adsorb.	ACFM	Required HP	Enrichmt.
alpha*	gamma	xf	x	F	y	Rf	xw/xf	% Removal	(lbmol/hr)	Area (ft^2)	(lbmol/min)	to Compressor	for Compr.	Ratio
200	0.2	0.001	0.000481	0.171028	0.00351558	0.213785	0.481	60.13	0.2759	83269	1.046	2500	334	3.516
200	0.2	0.001	0.000217	0.323999	0.00263366	0.404999	0.217	85.33	0.3916	157747	1.982	2500	334	2.634
200	0.2	0.001	0.000086	0.466748	0.00204422	0.583436	0.086	95.41	0.4378	227247	2.856	2500	334	2.044
20	0.2	0.001	0.000497	0.198114	0.00303594	0.247642	0.497	60.15	0.2760	9646	1.212	2500	334	3.036
20	0.2	0.001	0.000235	0.367020	0.00231935	0.458775	0.235	85.12	0.3906	17869	2.245	2500	334	2.319
20	0.2	0.001	0.000096	0.522897	0.00182482	0.653621	0.096	95.42	0.4378	25458	3.199	2500	334	1.825
5	0.2	0.001	0.000555	0.281933	0.00213338	0.352416	0.555	60.15	0.2760	3432	1.725	2500	334	2.133
5	0.2	0.001	0.00029	0.501576	0.00170553	0.626970	0.29	85.55	0.3925	6105	3.069	2500	334	1.706
5	0.2	0.001	0.00014	0.669099	0.00142530	0.836374	0.14	95.37	0.4376	8144	4.094	2500	334	1.425

TABLE F-1. (Continued)

C. Membrane Preconcentrator with Compressor and Carbon Adsorber

	FLOW (SCFM)	CONC (PPMV)
CASE 1	219.6	1000

alpha*	% Removal	SCFM to Carbon Adsorber	Reqd Amt of Carbon (pounds)	Superficial Velocity (ft/min)	Bed Area Normal to Flow (ft^2)	Bed Depth (ft)	new Ao (@ 3.0ft)	new Vs	Total Area (ft^2)
200	60.13	37.57	37	66	0.6	2.051	0.4	96.84	0.6
200	85.33	71.16	55	68	1.1	1.651	0.6	124.30	1.1
200	95.41	102.52	62	70	1.6	1.336	0.7	156.93	1.6
20	60.15	43.51	38	67	0.7	1.839	0.4	109.98	0.7
20	85.12	80.61	55	69	1.2	1.471	0.6	141.14	1.2
20	95.42	114.85	62	70	1.7	1.202	0.7	175.80	1.7
5	60.15	61.92	40	70	0.9	1.419	0.4	147.31	0.9
5	85.55	110.17	57	71	1.7	1.152	0.6	184.26	1.7
5	95.37	146.96	65	71	2.2	0.993	0.7	215.89	2.2

	FLOW (SCFM)	CONC (PPMV)
CASE 2	2196.4	1000

alpha*	% Removal	SCFM to Carbon Adsorber	Reqd Amt of Carbon (pounds)	Superficial Velocity (ft/min)	Bed Area Normal to Flow (ft^2)	Bed Depth (ft)	new Ao (@ 3.0ft)	new Vs	Total Area (ft^2)
200	60.13	375.65	370	66	6.0	2.051	4.1	96.84	6.0
200	85.33	711.64	547	68	11.0	1.651	6.1	124.30	11.0
200	95.41	1025.18	624	70	15.6	1.336	6.9	156.93	15.6
20	60.15	435.14	378	67	6.8	1.839	4.2	109.98	6.8
20	85.12	806.13	545	69	12.4	1.471	6.1	141.14	12.4
20	95.42	1148.51	624	70	17.3	1.202	6.9	175.80	17.3
5	60.15	619.25	401	70	9.4	1.419	4.5	147.31	9.4
5	85.55	1101.68	571	71	16.5	1.152	6.3	184.26	16.5
5	95.37	1469.63	650	71	21.8	0.993	7.2	215.89	21.8

TABLE F-1. (Continued)

CASE 3 8785.7 1000

Approx. Weller & Steiner Solution (xf-->0)

Constants				Calculated					Solvent Flux		Gas Flow to Adsorb.	ACFM	Required HP	Enrichmt.
alpha*	gamma	xf	x	F	y	Rf	xw/xf	% Removal	(lbmol/hr)	Area (ft^2)	(lbmol/min)	to Compressor	for Compr.	Ratio
200	0.2	0.001	0.000481	0.171028	0.00351558	0.213785	0.481	60.13	1.1036	333076	4.186	10000	1337	3.516
200	0.2	0.001	0.000217	0.323999	0.00263366	0.404999	0.217	85.33	1.5662	630986	7.929	10000	1337	2.634
200	0.2	0.001	0.000086	0.466748	0.00204422	0.583436	0.086	95.41	1.7513	908989	11.423	10000	1337	2.044
20	0.2	0.001	0.000497	0.198114	0.00303594	0.247642	0.497	60.15	1.1040	38583	4.848	10000	1337	3.036
20	0.2	0.001	0.000235	0.367020	0.00231935	0.458775	0.235	85.12	1.5624	71477	8.982	10000	1337	2.319
20	0.2	0.001	0.000096	0.522897	0.00182482	0.653621	0.096	95.42	1.7514	101834	12.797	10000	1337	1.825
5	0.2	0.001	0.000555	0.281933	0.00213338	0.352416	0.555	60.15	1.1040	13727	6.900	10000	1337	2.133
5	0.2	0.001	0.00029	0.501576	0.00170553	0.626970	0.29	85.55	1.5702	24420	12.275	10000	1337	1.706
5	0.2	0.001	0.00014	0.669099	0.00142530	0.836374	0.14	95.37	1.7504	32577	16.375	10000	1337	1.425

CASE 4 219.6 100

Approx. Weller & Steiner Solution (xf-->0)

Constants				Calculated					Solvent Flux		Gas Flow to Adsorb.	ACFM	Required HP	Enrichmt.
alpha*	gamma	xf	x	F	y	Rf	xw/xf	% Removal	(lbmol/hr)	Area (ft^2)	(lbmol/min)	to Compressor	for Compr.	Ratio
200	0.2	0.0001	0.000048	0.171028	0.00035155	0.213785	0.481	60.13	0.0028	8327	0.105	250	33	3.516
200	0.2	0.0001	0.000021	0.323999	0.00026336	0.404999	0.217	85.33	0.0039	15775	0.198	250	33	2.634
200	0.2	0.0001	0.000008	0.466748	0.00020442	0.583436	0.086	95.41	0.0044	22725	0.286	250	33	2.044
20	0.2	0.0001	0.000049	0.198114	0.00030359	0.247642	0.497	60.15	0.0028	965	0.121	250	33	3.036
20	0.2	0.0001	0.000023	0.367020	0.00023193	0.458775	0.235	85.12	0.0039	1787	0.225	250	33	2.319
20	0.2	0.0001	0.000009	0.522897	0.00018248	0.653621	0.096	95.42	0.0044	2546	0.320	250	33	1.825
5	0.2	0.0001	0.000055	0.281933	0.00021333	0.352416	0.555	60.15	0.0028	343	0.172	250	33	2.133
5	0.2	0.0001	0.000029	0.501576	0.00017055	0.626970	0.29	85.55	0.0039	611	0.307	250	33	1.706
5	0.2	0.0001	0.000014	0.669099	0.00014253	0.836374	0.14	95.37	0.0044	814	0.409	250	33	1.425

TABLE F-1. (Continued)

CASE 3 8785.7 1000

alpha*	% Removal	SCFM to Carbon Adsorbe	Reqd Amt of Carbon (pounds)	Superficial Velocity (ft/min)	Bed Area Normal to Flow (ft^2)	Bed Depth (ft)	new Ao (@ 3.0ft)	new Vs	Total Area (ft^2)
200	60.13	1502.61	1113	66	24.1	1.542	12.4	128.84	24.1
200	85.33	2846.57	1628	68	44.1	1.230	18.1	166.92	44.1
200	95.41	4100.72	1877	70	62.2	1.005	20.9	208.53	93.3
20	60.15	1740.57	1147	67	27.4	1.397	12.7	144.80	27.4
20	85.12	3224.53	1675	69	49.4	1.130	18.6	183.80	49.4
20	95.42	4594.03	3820	70	69.2	1.841	42.4	114.81	103.8
5	60.15	2476.99	1183	70	37.7	1.046	13.1	199.82	37.7
5	85.55	4406.70	1737	71	66.1	0.876	19.3	242.14	99.1
5	95.37	5878.52	1937	71	87.3	0.740	21.5	289.74	130.9

CASE 4 219.6 100

alpha*	% Removal	SCFM to Carbon Adsorber	Reqd Amt of Carbon (pounds)	Superficial Velocity (ft/min)	Bed Area Normal to Flow (ft^2)	Bed Depth (ft)	new Rt (@ 1.5ft)	new t	Total Area (ft^2)
200	60.13	37.57	5	96	0.4	0.397	18.7	9.45	0.4
200	85.33	71.16	7	97	0.8	0.309	34.9	12.12	0.8
200	95.41	102.52	8	98	1.1	0.256	49.7	14.65	1.1
20	60.15	43.51	5	97	0.5	0.345	21.5	10.86	0.5
20	85.12	80.61	7	98	0.9	0.281	39.3	13.33	0.9
20	95.42	114.85	9	99	1.2	0.236	55.5	15.90	1.2
5	60.15	61.92	5	98	0.7	0.267	30.1	14.06	0.7
5	85.55	110.17	8	99	1.2	0.227	53.2	16.50	1.2
5	95.37	146.96	9	99	1.6	0.197	70.6	19.08	1.6

TABLE F-1. (Continued)

CASE 5 2196.4 100

Approx. Weller & Steiner Solution (xf-->0)

Constants				Calculated					Solvent Flux		Gas Flow to Adsorb.	ACFM	Required HP	Enrichmt.
alpha*	gamma	xf	x	F	y	Rf	xw/xf	% Removal	(lbmol/hr)	Area (ft^2)	(lbmol/min)	to Compressor	for Compr.	Ratio
200	0.2	0.0001	0.000048	0.171028	0.00035155	0.213785	0.481	60.13	0.0276	83269	1.046	2500	334	3.516
200	0.2	0.0001	0.000021	0.323999	0.00026336	0.404999	0.217	85.33	0.0392	157747	1.982	2500	334	2.634
200	0.2	0.0001	0.000008	0.466748	0.00020442	0.583436	0.086	95.41	0.0438	227247	2.856	2500	334	2.044
20	0.2	0.0001	0.000049	0.198114	0.00030359	0.247642	0.497	60.15	0.0276	9646	1.212	2500	334	3.036
20	0.2	0.0001	0.000023	0.367020	0.00023193	0.458775	0.235	85.12	0.0391	17869	2.245	2500	334	2.319
20	0.2	0.0001	0.000009	0.522897	0.00018248	0.653621	0.096	95.42	0.0438	25458	3.199	2500	334	1.825
5	0.2	0.0001	0.000055	0.281933	0.00021333	0.352416	0.555	60.15	0.0276	3432	1.725	2500	334	2.133
5	0.2	0.0001	0.000029	0.501576	0.00017055	0.626970	0.29	85.55	0.0393	6105	3.069	2500	334	1.706
5	0.2	0.0001	0.000014	0.669099	0.00014253	0.836374	0.14	95.37	0.0438	8144	4.094	2500	334	1.425

CASE 6 8785.7 100

Approx. Weller & Steiner Solution (xf-->0)

Constants				Calculated					Solvent Flux		Gas Flow to Adsorb.	ACFM	Required HP	Enrichmt.
alpha*	gamma	xf	x	F	y	Rf	xw/xf	% Removal	(lbmol/hr)	Area (ft^2)	(lbmol/min)	to Compressor	for Compr.	Ratio
200	0.2	0.0001	0.000048	0.171028	0.00035155	0.213785	0.481	60.13	0.1104	333076	4.186	10000	1337	3.516
200	0.2	0.0001	0.000021	0.323999	0.00026336	0.404999	0.217	85.33	0.1566	630986	7.929	10000	1337	2.634
200	0.2	0.0001	0.000008	0.466748	0.00020442	0.583436	0.086	95.41	0.1751	908989	11.423	10000	1337	2.044
20	0.2	0.0001	0.000049	0.198114	0.00030359	0.247642	0.497	60.15	0.1104	38583	4.848	10000	1337	3.036
20	0.2	0.0001	0.000023	0.367020	0.00023193	0.458775	0.235	85.12	0.1562	71477	8.982	10000	1337	2.319
20	0.2	0.0001	0.000009	0.522897	0.00018248	0.653621	0.096	95.42	0.1751	101834	12.797	10000	1337	1.825
5	0.2	0.0001	0.000055	0.281933	0.00021333	0.352416	0.555	60.15	0.1104	13727	6.900	10000	1337	2.133
5	0.2	0.0001	0.000029	0.501576	0.00017055	0.626970	0.29	85.55	0.1570	24420	12.275	10000	1337	1.706
5	0.2	0.0001	0.000014	0.669099	0.00014253	0.836374	0.14	95.37	0.1750	32577	16.375	10000	1337	1.425

TABLE F-1. (Continued)

CASE 5 2196.4 100

alpha*	% Removal	SCFM to Carbon Adsorber	Reqd Amt of Carbon (pounds)	Superficial Velocity (ft/min)	Bed Area Normal to Flow (ft^2)	Bed Depth (ft)	new Rt (@ 1.5ft)	new t	Total Area (ft^2)
200	60.13	375.65	49	96	4.1	0.397	186.7	9.45	4.1
200	85.33	711.64	72	97	7.7	0.309	348.6	12.12	7.7
200	95.41	1025.18	85	98	11.1	0.256	497.5	14.65	11.1
20	60.15	435.14	49	97	4.8	0.345	214.6	10.86	4.8
20	85.12	806.13	74	98	8.7	0.281	392.9	13.33	8.7
20	95.42	1148.51	87	99	12.3	0.236	555.4	15.90	12.3
5	60.15	619.25	54	98	6.7	0.267	300.9	14.06	6.7
5	85.55	1101.68	81	99	11.8	0.227	531.7	16.50	11.8
5	95.37	1469.63	93	99	15.7	0.197	706.2	19.08	15.7

CASE 6 8785.7 100

alpha*	% Removal	SCFM to Carbon Adsorber	Reqd Amt of Carbon (pounds)	Superficial Velocity (ft/min)	Bed Area Normal to Flow (ft^2)	Bed Depth (ft)	new Rt (@ 1.5ft)	new t	Total Area (ft^2)
200	60.13	1502.61	135	96	16.6	0.272	746.8	13.81	16.6
200	85.33	2846.57	199	97	31.0	0.214	1394.6	17.52	31.0
200	95.41	4100.72	231	98	44.2	0.174	1990.0	21.53	66.3
20	60.15	1740.57	135	97	19.1	0.236	858.3	15.87	19.1
20	85.12	3224.53	206	98	34.9	0.197	1571.7	19.06	34.9
20	95.42	4594.03	240	99	49.4	0.162	2221.5	23.11	74.1
5	60.15	2476.99	151	98	26.7	0.189	1203.7	19.87	26.7
5	85.55	4406.70	224	99	47.3	0.158	2126.9	23.70	70.9
5	95.37	5878.52	261	99	62.8	0.139	2824.6	27.05	94.2

APPENDIX G

DETAILED COST ESTIMATES

TABLE G-1. SPREADSHEET CALCULATIONS FOR SYSTEM CAPITAL AND OPERATING COSTS

INPUT DATA

CASE No.	Inlet Flow (acfm)	Solvent Inlet Conc. (ppmv)	Solvent Name	Solvent Molec. Wt.	Inlet Temp. (deg F)	Inlet RH (%)	Total Inlet Molar Flow (lbmol/min)	Total Solvent Inlet Molar Flow (lbmol/min)	(lbmol/hr)
1	250	1000	CFC-113	187.4	100	18	0.6118	0.00061	0.037
2	2500	1000	CFC-113	187.4	100	18	6.1182	0.00612	0.367
3	10000	1000	Toluene	92.1	100	18	24.4727	0.02447	1.468
4	250	100	CFC-113	187.4	100	18	0.6118	0.00006	0.004
5	2500	100	CFC-113	187.4	100	18	6.1182	0.00061	0.037
6	10000	100	Toluene	92.1	100	18	24.4727	0.00245	0.147

A. BASE CASE: Direct Treatment using Carbon Adsorption

CARBON ADSORPTION	FLOW (SCFM)	CONC (PPMV)	DCCa (ads. vessels)	DCCd (duct work)	DCCc (carbon)	DCCf (fans)	TOTAL INSTALLED DIRECT CAPITAL	ANNUAL COST FOR CAPITAL RECOVERY	Blower HP	ACcb (carbon)	ACe (elect.)	Maint. Hours
CASE 1	219.6	1000	$45,000	$4,669	$322	$2,030	$52,021	$8,464	0.32	$99	$113	77.8
CASE 2	2196.4	1000	$152,554	$5,769	$3,220	$2,611	$164,154	$26,708	3.24	$992	$1,133	80.2
CASE 3	8785.7	1000	$303,512	$9,331	$19,321	$4,550	$336,714	$54,783	12.94	$5,954	$4,532	88.5
CASE 4	219.6	100	$45,000	$4,669	$242	$2,047	$51,958	$8,454	0.44	$74	$153	77.8
CASE 5	2196.4	100	$144,422	$5,769	$2,415	$2,785	$155,391	$25,282	4.36	$744	$1,526	80.2
CASE 6	8785.7	100	$260,561	$9,331	$14,490	$5,244	$289,626	$47,122	17.43	$4,466	$6,103	88.5

TABLE G-1. (Continued)

A. BASE CASE: Direct Treatment using Carbon Adsorption

ACm (Maint.)	ACol (Op. Labor)	Steam Required (lb/regen)	Solvent Recovered	ACp (credit for solvent)	ACs (steam)	ACcw (cooling water)	ACww (waste water)	TOTAL ANNUAL OPERATING COST	ANNUAL COST FOR CAPITAL RECOVERY	TOTAL ANNUAL COSTS	COST PER METRIC TON CONTROLLED
$3,708	$9,785	70	6.54	($3,353)	$643	$313	$939	$12,247	$8,464	$20,711	$849.42
$3,826	$9,785	699	65.35	($33,526)	$6,425	$3,131	$9,392	$1,158	$26,708	$27,866	$114.28
$4,218	$9,785	2795	128.47	($65,907)	$18,946	$9,232	$27,696	$14,456	$54,783	$69,239	$144.45
$3,708	$9,785	52	0.65	($335)	$161	$78	$235	$13,858	$8,454	$22,312	$9,150.55
$3,826	$9,785	524	6.54	($3,353)	$1,606	$783	$2,348	$17,265	$25,282	$42,547	$1,744.97
$4,218	$9,785	2097	12.85	($6,591)	$2,368	$1,154	$3,462	$24,966	$47,122	$72,088	$1,503.94

TABLE G-1. (Continued)

B. Membrane Preconcentrator with Vacuum Pump and Carbon Adsorber

	FLOW (SCFM)	CONC (PPMV)
CASE 1	219.6	1000

alpha*	% Removal	Membrane Area (ft^2)	ACFM to Vac. Pump	Req'd HP for Vacuum Pump	Installed Capital Cost	Membrane Capital Cost	Auxil. Equip. Cost	DCCa (ads. vessels)	DCCd (duct work)	DCCc (carbon)	DCCf (fans)	TOTAL ADSORBER DIRECT CAPITAL
200	60.02	20178	228	13	$39,246	$187,438	$93,719	$45,000	$4,546	$79	$1,965	$51,590
200	85.72	40649	460	26	$55,704	$377,602	$188,801	$45,000	$4,550	$115	$1,966	$51,631
200	95.68	61811	700	40	$68,690	$574,187	$287,094	$45,000	$4,549	$136	$1,965	$51,650
20	60.70	2797	317	18	$46,209	$25,985	$12,992	$45,000	$4,546	$81	$1,965	$51,593
20	85.22	5348	605	34	$63,892	$49,677	$24,839	$45,000	$4,549	$121	$1,9[illegible]	$51,636
20	95.54	8017	907	51	$78,230	$74,477	$37,238	$45,000	$4,548	$141	$1,965	$51,654
5	60.21	1256	569	32	$61,925	$11,667	$5,833	$45,000	$4,546	$89	$1,965	$51,600
5	85.42	2301	1042	59	$83,821	$21,375	$10,688	$45,000	$4,548	$131	$1,966	$51,644
5	95.34	3181	1440	82	$98,558	$29,552	$14,776	$45,000	$4,547	$150	$1,965	$51,662

	FLOW (SCFM)	CONC (PPMV)
CASE 2	2196.4	1000

alpha*	% Removal	Membrane Area (ft^2)	ACFM to Vac. Pump	Req'd HP for Vacuum Pump	Installed Capital Cost	Membrane Capital Cost	Auxil. Equip. Cost	DCCa (ads. vessels)	DCCd (duct work)	DCCc (carbon)	DCCf (fans)	TOTAL ADSORBER DIRECT CAPITAL
200	60.02	201777	2284	129	$124,107	$1,874,385	$937,192	$123,582	$4,546	$791	$1,965	$130,884
200	85.72	406488	4601	261	$176,150	$3,776,024	$1,888,012	$126,928	$4,550	$1,150	$1,966	$134,594
200	95.68	618111	6997	397	$217,217	$5,741,874	$2,870,937	$130,168	$4,549	$1,357	$1,965	$138,039
20	60.70	27973	3166	180	$146,125	$259,848	$129,924	$124,703	$4,546	$814	$1,965	$132,028
20	85.22	53477	6053	343	$202,043	$496,770	$248,385	$128,795	$4,549	$1,209	$1,966	$136,519
20	95.54	80174	9075	514	$247,386	$744,765	$372,383	$132,875	$4,548	$1,410	$1,965	$140,798
5	60.21	12559	5686	322	$195,825	$116,666	$58,333	$127,973	$4,546	$888	$1,965	$135,373
5	85.42	23010	10418	591	$265,065	$213,753	$106,877	$134,501	$4,548	$1,313	$1,966	$142,327
5	95.34	31813	14404	817	$311,668	$295,523	$147,761	$139,837	$4,547	$1,497	$1,965	$147,846

TABLE G-1. (Continued)

B. Membrane Preconcentrator with Vacuum Pump and Carbon Adsorber

CASE 1 219.6 1000

TOTAL INSTALLED DIRECT CAPITAL	ANNUAL COST FOR CAPITAL RECOVERY	Blower HP	ACcb (carbon)	ACe (elect.)	Maint. Hours	ACm (Maint.)	ACol (Op. Labor)	Steam Required (lb/regen)	Solvent Recovered	ACp (credit for solvent)	ACs (steam)	ACmem (membrane replacement)	ACcw (cooling water)
$371,994	$60,523	0.04	$24	$4,548	77.5	$3,696	$9,785	17	3.84	($1,972)	$310	$62,479	$151
$673,738	$109,617	0.08	$35	$9,159	77.6	$3,697	$9,785	25	5.49	($2,816)	$451	$125,867	$220
$981,621	$159,710	0.10	$42	$13,923	77.6	$3,698	$9,785	29	6.13	($3,144)	$532	$191,396	$259
$136,779	$22,254	0.05	$25	$6,303	77.5	$3,696	$9,785	18	3.89	($1,994)	$319	$8,662	$155
$190,043	$30,920	0.09	$37	$12,046	77.6	$3,698	$9,785	26	5.46	($2,800)	$474	$16,559	$231
$241,599	$39,308	0.12	$43	$18,053	77.6	$3,700	$9,785	31	6.12	($3,139)	$552	$24,826	$269
$131,025	$21,318	0.07	$27	$11,312	77.6	$3,698	$9,785	19	3.86	($1,978)	$348	$3,889	$170
$167,529	$27,257	0.12	$40	$20,719	77.6	$3,700	$9,785	28	5.47	($2,806)	$515	$7,125	$251
$194,548	$31,653	0.14	$46	$28,638	77.7	$3,702	$9,785	32	6.11	($3,132)	$586	$9,851	$286

CASE 2 2196.4 1000

TOTAL INSTALLED DIRECT CAPITAL	ANNUAL COST FOR CAPITAL RECOVERY	Blower HP	ACcb (carbon)	ACe (elect.)	Maint. Hours	ACm (Maint.)	ACol (Op. Labor)	Steam Required (lb/regen)	Solvent Recovered	ACp (credit for solvent)	ACs (steam)	ACmem (membrane replacement)	ACcw (cooling water)
$3,066,568	$498,931	0.44	$244	$45,485	77.8	$3,707	$9,785	172	38.44	($19,720)	$3,099	$624,795	$1,510
$5,974,780	$972,097	0.77	$354	$91,591	78.0	$3,719	$9,785	250	54.90	($28,164)	$4,506	$1,258,675	$2,196
$8,968,067	$1,459,104	1.02	$418	$139,225	78.3	$3,731	$9,785	295	61.28	($31,436)	$5,320	$1,913,958	$2,592
$667,925	$108,671	0.54	$251	$63,031	77.8	$3,711	$9,785	177	38.88	($19,943)	$3,191	$86,616	$1,555
$1,083,718	$176,321	0.90	$373	$120,459	78.2	$3,727	$9,785	262	54.58	($28,000)	$4,739	$165,590	$2,309
$1,505,332	$244,917	1.16	$434	$180,527	78.5	$3,742	$9,785	306	61.19	($31,391)	$5,525	$248,255	$2,692
$506,197	$82,358	0.73	$274	$113,118	78.1	$3,725	$9,785	193	38.56	($19,784)	$3,482	$38,889	$1,697
$728,023	$118,449	1.17	$405	$207,192	78.6	$3,749	$9,785	285	54.71	($28,065)	$5,145	$71,251	$2,507
$902,798	$146,885	1.42	$461	$286,383	79.1	$3,770	$9,785	325	61.06	($31,324)	$5,865	$98,508	$2,858

TABLE G-1. (Continued)

B. Membrane Preconcentrator with Vacuum Pump and Carbon Adsor

CASE 1 219.6 1000

ACww (waste water)	TOTAL ANNUAL OPERATING COST	ANNUAL COST FOR CAPITAL RECOVERY	TOTAL ANNUAL COSTS	COST PER METRIC TON CONTROLLED
$453	$79,475	$60,523	$139,998	$9,762
$659	$147,056	$109,617	$256,674	$12,531
$778	$217,269	$159,710	$376,978	$16,489
$466	$27,418	$22,254	$49,672	$3,425
$693	$40,722	$30,920	$71,642	$3,518
$808	$54,896	$39,308	$94,204	$4,126
$509	$27,759	$21,318	$49,077	$3,411
$752	$40,081	$27,257	$67,337	$3,299
$857	$50,619	$31,653	$82,272	$3,611

CASE 2 2196.4 1000

ACww (waste water)	TOTAL ANNUAL OPERATING COST	ANNUAL COST FOR CAPITAL RECOVERY	TOTAL ANNUAL COSTS	COST PER METRIC TON CONTROLLED
$4,530	$673,434	$498,931	$1,172,364	$8,174
$6,587	$1,349,249	$972,097	$2,321,346	$11,333
$7,777	$2,051,371	$1,459,104	$3,510,475	$15,355
$4,665	$152,861	$108,671	$261,533	$1,803
$6,927	$285,907	$176,321	$462,228	$2,270
$8,076	$427,646	$244,917	$672,563	$2,946
$5,090	$156,274	$82,358	$238,633	$1,659
$7,521	$279,491	$118,449	$397,940	$1,950
$8,573	$384,879	$146,885	$531,764	$2,334

TABLE G-1. (Continued)

CASE 3 8785.7 1000

alpha*	% Removal	Membrane Area (ft^2)	ACFM to Vac. Pump	Req'd HP for Vacuum Pump	Installed Capital Cost	Membrane Capital Cost	Auxil. Equip. Cost	DCCa (ads. vessels)	DCCd (duct work)	DCCc (carbon)	DCCf (fans)	TOTAL ADSORBER DIRECT CAPITAL
200	60.02	807107	9136	518	$248,214	$7,497,539	$3,748,770	$135,179	$4,546	$2,353	$1,965	$144,043
200	85.72	1625950	18404	1043	$352,301	$15,104,097	$7,552,049	$148,380	$4,550	$3,552	$1,966	$158,448
200	95.68	2472442	27986	1587	$434,433	$22,967,497	$11,483,748	$161,062	$4,549	$4,081	$1,965	$171,657
20	60.70	111890	12665	718	$292,251	$1,039,392	$519,696	$139,612	$4,546	$2,515	$1,955	$148,638
20	85.22	213909	24213	1373	$404,086	$1,987,081	$993,541	$155,702	$4,549	$3,635	$1,966	$165,852
20	95.54	320695	36300	2058	$494,773	$2,979,060	$1,489,530	$171,577	$4,548	$4,199	$1,965	$182,289
5	60.21	50236	22745	1289	$391,650	$466,664	$233,332	$152,485	$4,546	$2,646	$1,965	$161,642
5	85.42	92042	41674	2363	$530,130	$855,014	$427,507	$177,859	$4,548	$3,871	$1,966	$188,244
5	95.34	127252	57615	3266	$623,335	$1,182,092	$591,046	$235,595	$4,547	$6,690	$1,965	$248,798

CASE 4 219.6 100

alpha*	% Removal	Membrane Area (ft^2)	ACFM to Vac. Pump	Req'd HP for Vacuum Pump	Installed Capital Cost	Membrane Capital Cost	Auxil. Equip. Cost	DCCa (ads. vessels)	DCCd (duct work)	DCCc (carbon)	DCCf (fans)	TOTAL ADSORBER DIRECT CAPITAL
200	60.02	20178	228	13	$39,246	$187,438	$93,719	$45,000	$4,546	$24	$1,965	$51,535
200	85.72	40649	460	26	$55,704	$377,602	$188,801	$45,000	$4,550	$47	$1,967	$51,564
200	95.68	61811	700	40	$68,690	$574,187	$287,094	$45,000	$4,549	$70	$1,965	$51,584
20	60.70	2797	317	18	$46,209	$25,985	$12,992	$45,000	$4,546	$33	$1,965	$51,544
20	85.22	5348	605	34	$63,892	$49,677	$24,839	$45,000	$4,549	$61	$1,966	$51,576
20	95.54	8017	907	51	$78,230	$74,477	$37,238	$45,000	$4,548	$90	$1,965	$51,603
5	60.21	1256	569	32	$61,925	$11,667	$5,833	$45,000	$4,546	$56	$1,965	$51,568
5	85.42	2301	1042	59	$83,821	$21,375	$10,688	$45,000	$4,548	$102	$1,966	$51,616
5	95.34	3181	1440	82	$98,558	$29,552	$14,776	$45,000	$4,547	$141	$1,965	$51,653

TABLE G-1. (Continued)

CASE 3 8785.7 1000

TOTAL INSTALLED DIRECT CAPITAL	ANNUAL COST FOR CAPITAL RECOVERY	Blower HP	ACcb (carbon)	ACe (elect.)	Maint. Hours	ACm (Maint.)	ACol (Op. Labor)	Steam Required (lb/regen)	Solvent Recovered	ACp (credit for solvent)	ACs (steam)	ACmem (membrane replacement)	ACcw (cooling water)
$11,638,565	$1,893,594	1.35	$725	$181,800	78.5	$3,743	$9,785	511	75.57	($38,767)	$9,220	$2,499,180	$4,493
$23,166,895	$3,769,254	2.46	$1,095	$366,151	79.5	$3,791	$9,785	771	107.93	($55,367)	$13,921	$5,034,699	$6,783
$35,057,335	$5,703,828	3.23	$1,258	$556,596	80.6	$3,841	$9,785	886	120.47	($61,799)	$15,995	$7,655,832	$7,794
$1,999,978	$325,396	1.72	$775	$251,977	78.9	$3,761	$9,785	546	76.42	($39,206)	$9,857	$346,464	$4,803
$3,550,560	$577,676	2.84	$1,120	$481,568	80.2	$3,822	$9,785	789	107.30	($55,044)	$14,247	$662,360	$6,942
$5,145,652	$837,198	3.66	$1,294	$721,763	81.5	$3,885	$9,785	911	120.29	($61,710)	$16,456	$993,020	$8,019
$1,253,288	$203,910	2.30	$816	$452,253	80.0	$3,814	$9,785	574	75.81	($38,892)	$10,371	$155,555	$5,054
$2,000,895	$325,546	3.67	$1,193	$828,422	82.1	$3,913	$9,785	840	107.55	($55,172)	$15,172	$285,005	$7,393
$2,645,271	$430,386	4.54	$2,062	$1,145,137	83.8	$3,996	$9,785	1452	120.04	($61,578)	$26,220	$394,031	$12,777

CASE 4 219.6 100

TOTAL INSTALLED DIRECT CAPITAL	ANNUAL COST FOR CAPITAL RECOVERY	Blower HP	ACcb (carbon)	ACe (elect.)	Maint. Hours	ACm (Maint.)	ACol (Op. Labor)	Steam Required (lb/regen)	Solvent Recovered	ACp (credit for solvent)	ACs (steam)	ACmem (membrane replacement)	ACcw (cooling water)
$371,939	$60,514	0.02	$7	$4,540	77.5	$3,696	$9,785	5	0.38	($197)	$41	$62,479	$20
$673,671	$109,606	0.03	$15	$9,143	77.6	$3,697	$9,785	10	0.55	($282)	$62	$125,867	$30
$981,555	$159,699	0.04	$22	$13,901	77.6	$3,698	$9,785	15	0.61	($314)	$73	$191,396	$35
$136,730	$22,246	0.02	$10	$6,292	77.5	$3,696	$9,785	7	0.39	($199)	$44	$8,662	$21
$189,983	$30,910	0.04	$19	$12,027	77.6	$3,698	$9,785	13	0.55	($280)	$65	$16,559	$32
$241,548	$39,300	0.05	$28	$18,029	77.6	$3,700	$9,785	20	0.61	($314)	$75	$24,826	$36
$130,993	$21,313	0.03	$17	$11,297	77.6	$3,698	$9,785	12	0.39	($198)	$47	$3,889	$23
$167,500	$27,252	0.05	$32	$20,696	77.6	$3,700	$9,785	22	0.55	($281)	$71	$7,125	$34
$194,539	$31,652	0.06	$43	$28,611	77.7	$3,702	$9,785	31	0.61	($313)	$81	$9,851	$40

TABLE G-1. (Continued)

CASE 3 8785.7 1000

ACww (waste water)	TOTAL ANNUAL OPERATING COST	ANNUAL COST FOR CAPITAL RECOVERY	TOTAL ANNUAL COSTS	COST PER METRIC TON CONTROLLED
$13,478	$2,683,656	$1,893,594	$4,577,251	$16,235
$20,350	$5,401,208	$3,769,254	$9,170,462	$22,774
$23,382	$8,212,684	$5,703,828	$13,916,513	$30,964
$14,410	$602,626	$325,396	$928,023	$3,255
$20,826	$1,145,626	$577,676	$1,723,302	$4,305
$24,056	$1,716,567	$837,198	$2,553,764	$5,690
$15,161	$613,916	$203,910	$817,826	$2,891
$22,179	$1,117,890	$325,546	$1,443,435	$3,597
$38,330	$1,570,759	$430,386	$2,001,144	$4,468

CASE 4 219.6 100

ACww (waste water)	TOTAL ANNUAL OPERATING COST	ANNUAL COST FOR CAPITAL RECOVERY	TOTAL ANNUAL COSTS	COST PER METRIC TON CONTROLLED
$60	$80,432	$60,514	$140,946	$98,276
$91	$148,408	$109,606	$258,015	$125,964
$106	$218,702	$159,699	$378,401	$165,510
$64	$28,375	$22,246	$50,621	$34,901
$95	$41,999	$30,910	$72,909	$35,803
$109	$56,272	$39,300	$95,572	$41,863
$69	$28,626	$21,313	$49,939	$34,708
$103	$41,265	$27,252	$68,517	$33,569
$119	$51,918	$31,652	$83,569	$36,683

TABLE G-1. (Continued)

CASE 5 2196.4 100

alpha*	% Removal	Membrane Area (ft^2)	ACFM to Vac. Pump	Req'd HP for Vacuum Pump	Installed Capital Cost	Membrane Capital Cost	Auxil. Equip. Cost	DCCa (ads. vessels)	DCCd (duct work)	DCCc (carbon)	DCCf (fans)	TOTAL ADSORBER DIRECT CAPITAL
200	60.02	201777	2284	129	$124,107	$1,874,385	$937,192	$122,192	$4,546	$242	$1,965	$128,945
200	85.72	406488	4601	261	$176,150	$3,776,024	$1,888,012	$124,562	$4,550	$472	$1,967	$131,550
200	95.68	618111	6997	397	$217,217	$5,741,874	$2,870,937	$126,940	$4,549	$703	$1,965	$134,156
20	60.70	27973	3166	180	$146,125	$259,848	$129,924	$123,053	$4,546	$325	$1,965	$129,890
20	85.22	53477	6053	343	$202,043	$496,770	$248,385	$125,975	$4,549	$609	$1,966	$133,099
20	95.54	80174	9075	514	$247,386	$744,765	$372,383	$128,965	$4,548	$900	$1,965	$136,378
5	60.21	12559	5686	322	$195,825	$116,666	$58,333	$125,512	$4,546	$564	$1,965	$132,587
5	85.42	23010	10418	591	$265,065	$213,753	$106,877	$130,233	$4,548	$1,023	$1,966	$137,770
5	95.34	31813	14404	817	$311,668	$295,523	$147,761	$134,150	$4,547	$1,406	$1,965	$142,068

CASE 6 8785.7 100

alpha*	% Removal	Membrane Area (ft^2)	ACFM to Vac. Pump	Req'd HP for Vacuum Pump	Installed Capital Cost	Membrane Capital Cost	Auxil. Equip. Cost	DCCa (ads. vessels)	DCCd (duct work)	DCCc (carbon)	DCCf (fans)	TOTAL ADSORBER DIRECT CAPITAL
200	60.02	807107	9136	518	$248,214	$7,497,539	$3,748,770	$129,661	$4,546	$968	$1,965	$137,140
200	85.72	1625950	18404	1043	$352,301	$15,104,097	$7,552,049	$139,056	$4,550	$1,887	$1,967	$147,460
200	95.68	2472442	27986	1587	$434,433	$22,967,497	$11,483,748	$148,427	$4,549	$2,811	$1,965	$157,752
20	60.70	111890	12665	718	$292,251	$1,039,392	$519,696	$133,081	$4,546	$1,301	$1,965	$140,893
20	85.22	213909	24213	1373	$404,086	$1,987,081	$993,541	$144,633	$4,549	$2,436	$1,966	$153,584
20	95.54	320695	36300	2058	$494,773	$2,979,060	$1,489,530	$156,367	$4,548	$3,599	$1,965	$166,479
5	60.21	50236	22745	1289	$391,650	$466,664	$233,332	$142,808	$4,546	$2,256	$1,965	$151,575
5	85.42	92042	41674	2363	$530,130	$855,014	$427,507	$161,315	$4,548	$4,094	$1,966	$171,922
5	95.34	127252	57615	3266	$623,335	$1,182,092	$591,046	$203,894	$4,547	$8,436	$1,965	$218,842

TABLE G-1. (Continued)

CASE 5 2196.4 100

TOTAL INSTALLED DIRECT CAPITAL	ANNUAL COST FOR CAPITAL RECOVERY	Blower HP	ACcb (carbon)	ACe (elect.)	Maint. Hours	ACm (Maint.)	ACol (Op. Labor)	Steam Required (lb/regen)	Solvent Recovered	ACp (credit for solvent)	ACs (steam)	ACmem (membrane replacement)	ACcw (cooling water)
$3,064,629	$498,615	0.18	$75	$45,396	77.8	$3,707	$9,785	53	3.84	($1,972)	$413	$624,795	$201
$5,971,737	$971,602	0.32	$145	$91,434	78.0	$3,719	$9,785	102	5.49	($2,816)	$620	$1,258,675	$302
$8,964,184	$1,458,473	0.42	$217	$139,013	78.3	$3,731	$9,785	153	6.13	($3,144)	$728	$1,913,958	$355
$665,787	$108,324	0.22	$100	$62,922	77.8	$3,711	$9,785	71	3.89	($1,994)	$439	$86,616	$214
$1,080,298	$175,764	0.37	$188	$120,272	78.2	$3,727	$9,785	132	5.46	($2,800)	$648	$165,590	$316
$1,500,912	$244,198	0.48	$277	$180,288	78.5	$3,742	$9,785	195	6.12	($3,139)	$747	$248,255	$364
$503,411	$81,905	0.30	$174	$112,967	78.1	$3,725	$9,785	122	3.86	($1,978)	$471	$38,889	$229
$723,465	$117,708	0.50	$315	$206,958	78.6	$3,749	$9,785	222	5.47	($2,806)	$706	$71,251	$344
$897,020	$145,945	0.63	$433	$286,107	79.1	$3,770	$9,785	305	6.11	($3,132)	$811	$98,508	$395

CASE 6 8785.7 100

TOTAL INSTALLED DIRECT CAPITAL	ANNUAL COST FOR CAPITAL RECOVERY	Blower HP	ACcb (carbon)	ACe (elect.)	Maint. Hours	ACm (Maint.)	ACol (Op. Labor)	Steam Required (lb/regen)	Solvent Recovered	ACp (credit for solvent)	ACs (steam)	ACmem (membrane replacement)	ACcw (cooling water)
$11,631,662	$1,892,471	0.58	$298	$181,531	78.5	$3,743	$9,785	210	7.56	($3,877)	$1,176	$2,499,180	$573
$23,155,906	$3,767,466	1.01	$582	$365,643	79.5	$3,791	$9,785	410	10.79	($5,537)	$1,740	$5,034,699	$848
$35,043,430	$5,701,566	1.34	$866	$555,936	80.6	$3,841	$9,785	610	12.05	($6,180)	$2,014	$7,655,832	$981
$1,992,233	$324,136	0.70	$401	$251,622	78.9	$3,761	$9,785	282	7.64	($3,921)	$1,232	$346,464	$600
$3,538,293	$575,680	1.18	$751	$480,985	80.2	$3,822	$9,785	529	10.73	($5,504)	$1,794	$662,360	$874
$5,129,842	$834,625	1.54	$1,109	$721,021	81.5	$3,885	$9,785	781	12.03	($6,171)	$2,011	$993,020	$980
$1,243,220	$202,272	0.97	$695	$451,787	80.0	$3,814	$9,785	490	7.58	($3,889)	$1,268	$155,555	$618
$1,984,573	$322,890	1.62	$1,262	$827,701	82.1	$3,913	$9,785	888	10.75	($5,517)	$1,867	$285,005	$910
$2,615,315	$425,512	2.08	$2,600	$1,144,278	83.8	$3,996	$9,785	1831	12.00	($6,158)	$3,251	$394,031	$1,584

TABLE G-1. (Continued)

CASE 5	2196.4	100

ACww (waste water)	TOTAL ANNUAL OPERATING COST	ANNUAL COST FOR CAPITAL RECOVERY	TOTAL ANNUAL COSTS	COST PER METRIC TON CONTROLLED
$604	$683,004	$498,615	$1,181,619	$82,390
$906	$1,362,768	$971,602	$2,334,370	$113,965
$1,064	$2,065,707	$1,458,473	$3,524,180	$154,145
$641	$162,434	$108,324	$270,757	$18,667
$948	$298,673	$175,764	$474,438	$23,298
$1,091	$441,410	$244,198	$685,608	$30,031
$688	$164,948	$81,905	$246,853	$17,157
$1,032	$291,333	$117,708	$409,041	$20,040
$1,185	$397,861	$145,945	$543,806	$23,871

CASE 6	8785.7	100

ACww (waste water)	TOTAL ANNUAL OPERATING COST	ANNUAL COST FOR CAPITAL RECOVERY	TOTAL ANNUAL COSTS	COST PER METRIC TON CONTROLLED
$1,720	$2,694,128	$1,892,471	$4,586,600	$162,681
$2,544	$5,414,095	$3,767,466	$9,181,561	$228,018
$2,944	$8,226,020	$5,701,566	$13,927,586	$309,883
$1,801	$611,746	$324,136	$935,882	$32,823
$2,623	$1,157,489	$575,680	$1,733,169	$43,294
$2,940	$1,728,580	$834,625	$2,563,206	$57,113
$1,853	$621,484	$202,272	$823,756	$29,123
$2,730	$1,127,655	$322,890	$1,450,545	$36,151
$4,753	$1,558,120	$425,512	$1,983,632	$44,293

TABLE G-1. (Continued)

C. Membrane Preconcentrator with Compressor and Carbon Adsorber

CASE 1 219.6 1000

alpha*	% Removal	Membrane Area (ft^2)	ACFM to Compr.	Req'd HP for Compr.	Installed Capital Cost	Membrane Capital Cost	Auxil. Equip. Cost	DCCa (ads. vessels)	DCCd (duct work)	DCCc (carbon)	DCCf (fans)	TOTAL ADSORBER DIRECT CAPITAL
200	60.13	8327	250	33	$30,000	$77,352	$38,676	$45,000	$4,546	$85	$1,965	$51,596
200	85.33	15775	250	33	$30,000	$146,537	$73,269	$45,000	$4,548	$126	$1,966	$51,640
200	95.41	22725	250	33	$30,000	$211,099	$105,550	$45,000	$4,548	$144	$1,965	$51,656
20	60.15	965	250	33	$30,000	$8,960	$4,480	$45,000	$4,546	$87	$1,965	$51,598
20	85.12	1787	250	33	$30,000	$16,599	$8,300	$45,000	$4,548	$126	$1,966	$51,639
20	95.42	2546	250	33	$30,000	$23,649	$11,825	$45,000	$4,547	$144	$1,965	$51,656
5	60.15	343	250	33	$30,000	$3,188	$1,594	$45,000	$4,546	$92	$1,965	$51,604
5	85.55	611	250	33	$30,000	$5,671	$2,836	$45,000	$4,547	$131	$1,965	$51,644
5	95.37	814	250	33	$30,000	$7,565	$3,783	$45,000	$4,547	$150	$1,965	$51,662

CASE 2 2196.4 1000

alpha*	% Removal	Membrane Area (ft^2)	ACFM to Compr.	Req'd HP for Compr.	Installed Capital Cost	Membrane Capital Cost	Auxil. Equip. Cost	DCCa (ads. vessels)	DCCd (duct work)	DCCc (carbon)	DCCf (fans)	TOTAL ADSORBER DIRECT CAPITAL
200	60.13	83269	2500	334	$128,000	$773,519	$386,760	$126,128	$4,546	$853	$1,965	$133,492
200	85.33	157747	2500	334	$128,000	$1,465,371	$732,685	$131,466	$4,548	$1,259	$1,966	$139,239
200	95.41	227247	2500	334	$128,000	$2,110,991	$1,055,495	$136,263	$4,548	$1,437	$1,965	$144,212
20	60.15	9646	2500	334	$128,000	$89,602	$44,801	$127,012	$4,546	$870	$1,965	$134,393
20	85.12	17869	2500	334	$128,000	$165,994	$82,997	$132,870	$4,548	$1,256	$1,966	$140,640
20	95.42	25458	2500	334	$128,000	$236,494	$118,247	$138,098	$4,547	$1,427	$1,965	$146,047
5	60.15	3432	2500	334	$128,000	$31,878	$15,939	$129,760	$4,546	$924	$1,965	$137,196
5	85.55	6105	2500	334	$128,000	$56,713	$28,356	$137,279	$4,547	$1,315	$1,965	$145,107
5	95.37	8144	2500	334	$128,000	$75,654	$37,827	$142,870	$4,547	$1,497	$1,965	$150,879

TABLE G-1. (Continued)

C. Membrane Preconcentrator with Compressor and Carbon Adsorber

CASE 1 219.6 1000

TOTAL INSTALLED DIRECT CAPITAL	ANNUAL COST FOR CAPITAL RECOVERY	Blower HP	ACcb (carbon)	ACe (elect.)	Maint. Hours	ACm (Maint.)	ACol (Op. Labor)	Steam Required (lb/regen)	Solvent Recovered	ACp (credit for solvent)	ACs (steam)	ACmem (membrane replacement)	ACcw (cooling water)
$197,624	$32,153	0.06	$26	$11,728	77.5	$3,697	$9,785	19	3.85	($1,976)	$334	$25,784	$163
$301,445	$49,045	0.10	$39	$11,742	77.6	$3,699	$9,785	27	5.47	($2,804)	$493	$48,846	$240
$398,305	$64,804	0.13	$44	$11,750	77.6	$3,701	$9,785	31	6.11	($3,135)	$563	$70,366	$274
$95,038	$15,463	0.07	$27	$11,729	77.6	$3,697	$9,785	19	3.85	($1,976)	$341	$2,987	$166
$106,538	$17,334	0.11	$39	$11,743	77.6	$3,700	$9,785	27	5.45	($2,797)	$492	$5,533	$240
$117,130	$19,057	0.13	$44	$11,752	77.6	$3,702	$9,785	31	6.11	($3,135)	$563	$7,883	$274
$86,385	$14,055	0.08	$28	$11,734	77.6	$3,699	$9,785	20	3.85	($1,976)	$362	$1,063	$177
$90,151	$14,668	0.12	$41	$11,749	77.6	$3,701	$9,785	29	5.48	($2,811)	$515	$1,890	$251
$93,010	$15,133	0.15	$46	$11,757	77.7	$3,704	$9,785	32	6.11	($3,133)	$587	$2,522	$286

CASE 2 2196.4 1000

TOTAL INSTALLED DIRECT CAPITAL	ANNUAL COST FOR CAPITAL RECOVERY	Blower HP	ACcb (carbon)	ACe (elect.)	Maint. Hours	ACm (Maint.)	ACol (Op. Labor)	Steam Required (lb/regen)	Solvent Recovered	ACp (credit for solvent)	ACs (steam)	ACmem (membrane replacement)	ACcw (cooling water)
$1,421,771	$231,322	0.64	$263	$117,276	78.0	$3,717	$9,785	185	38.51	($19,755)	$3,343	$257,840	$1,629
$2,465,295	$401,104	1.04	$388	$117,416	78.4	$3,737	$9,785	273	54.65	($28,036)	$4,934	$488,457	$2,404
$3,438,699	$559,476	1.28	$443	$117,502	78.8	$3,756	$9,785	312	61.11	($31,349)	$5,630	$703,664	$2,743
$396,797	$64,559	0.68	$268	$117,292	78.0	$3,721	$9,785	189	38.52	($19,762)	$3,410	$29,867	$1,662
$517,631	$84,219	1.08	$387	$117,431	78.5	$3,743	$9,785	273	54.52	($27,969)	$4,922	$55,331	$2,399
$628,788	$102,304	1.33	$443	$117,518	78.9	$3,763	$9,785	312	61.11	($31,351)	$5,630	$78,831	$2,744
$313,013	$50,927	0.81	$285	$117,337	78.3	$3,732	$9,785	201	38.52	($19,762)	$3,623	$10,626	$1,765
$358,176	$58,275	1.24	$405	$117,486	78.9	$3,760	$9,785	285	54.79	($28,107)	$5,153	$18,904	$2,511
$392,360	$63,837	1.49	$461	$117,573	79.3	$3,782	$9,785	325	61.08	($31,334)	$5,867	$25,218	$2,859

TABLE G-1. (Continued)

C. Membrane Preconcentrator with Compressor and Carbon Adsorb

CASE 1 219.6 1000

ACww (waste water)	TOTAL ANNUAL OPERATING COST	ANNUAL COST FOR CAPITAL RECOVERY	TOTAL ANNUAL COSTS	COST PER METRIC TON CONTROLLED
$489	$50,030	$32,153	$82,183	$5,720
$721	$72,761	$49,045	$121,806	$5,974
$823	$94,172	$64,804	$158,976	$6,973
$498	$27,254	$15,463	$42,717	$2,972
$720	$29,454	$17,334	$46,788	$2,300
$823	$31,691	$19,057	$50,748	$2,226
$530	$25,400	$14,055	$39,455	$2,745
$753	$25,875	$14,668	$40,542	$1,983
$858	$26,410	$15,133	$41,543	$1,823

CASE 2 2196.4 1000

ACww (waste water)	TOTAL ANNUAL OPERATING COST	ANNUAL COST FOR CAPITAL RECOVERY	TOTAL ANNUAL COSTS	COST PER METRIC TON CONTROLLED
$4,887	$378,984	$231,322	$610,306	$4,248
$7,213	$606,298	$401,104	$1,007,402	$4,941
$8,230	$820,403	$559,476	$1,379,879	$6,052
$4,985	$151,227	$64,559	$215,786	$1,501
$7,196	$173,225	$84,219	$257,444	$1,266
$8,231	$195,594	$102,304	$297,897	$1,307
$5,296	$132,686	$50,927	$183,614	$1,278
$7,533	$137,430	$58,275	$195,706	$957
$8,576	$142,787	$63,837	$206,624	$907

TABLE G-1. (Continued)

CASE 3 8785.7 1000

alpha*	% Removal	Membrane Area (ft^2)	ACFM to Compr.	Req'd HP for Compr.	Installed Capital Cost	Membrane Capital Cost	Auxil. Equip. Cost	DCCa (ads. vessels)	DCCd (duct work)	DCCc (carbon)	DCCf (fans)	TOTAL ADSORBER DIRECT CAPITAL
200	60.13	333076	10000	1337	$263,000	$3,094,077	$1,547,038	$145,235	$4,546	$2,565	$1,965	$154,311
200	85.33	630986	10000	1337	$263,000	$5,861,483	$2,930,741	$166,113	$4,548	$3,750	$1,966	$176,378
200	95.41	908989	10000	1337	$263,000	$8,443,964	$4,221,982	$215,769	$4,548	$6,487	$1,965	$[illegible]
20	60.15	38583	10000	1337	$263,000	$358,409	$179,204	$148,712	$4,546	$2,643	$1,965	$157,867
20	85.12	71477	10000	1337	$263,000	$663,977	$331,988	$171,560	$4,548	$3,858	$1,966	$181,931
20	95.42	101834	10000	1337	$263,000	$945,974	$472,987	$225,990	$4,547	$13,199	$1,965	$245,702
5	60.15	13727	10000	1337	$263,000	$127,512	$63,756	$159,473	$4,546	$2,726	$1,965	$168,710
5	85.55	24420	10000	1337	$263,000	$226,851	$113,425	$221,438	$4,547	$6,003	$1,965	$233,954
5	95.37	32577	10000	1337	$263,000	$302,617	$151,309	$252,169	$4,547	$6,693	$1,965	$265,374

CASE 4 219.6 100

alpha*	% Removal	Membrane Area (ft^2)	ACFM to Compr.	Req'd HP for Compr.	Installed Capital Cost	Membrane Capital Cost	Auxil. Equip. Cost	DCCa (ads. vessels)	DCCd (duct work)	DCCc (carbon)	DCCf (fans)	TOTAL ADSORBER DIRECT CAPITAL
200	60.13	8327	250	33	$30,000	$77,352	$38,676	$45,000	$4,546	$43	$1,965	$51,554
200	85.33	15775	250	33	$30,000	$146,537	$73,269	$45,000	$4,548	$80	$1,966	$51,594
200	95.41	22725	250	33	$30,000	$211,099	$105,550	$45,000	$4,548	$115	$1,965	$51,627
20	60.15	965	250	33	$30,000	$8,960	$4,480	$45,000	$4,546	$49	$1,965	$51,561
20	85.12	1787	250	33	$30,000	$16,599	$8,300	$45,000	$4,548	$91	$1,966	$51,604
20	95.42	2546	250	33	$30,000	$23,649	$11,825	$45,000	$4,547	$128	$1,965	$51,640
5	60.15	343	250	33	$30,000	$3,188	$1,594	$45,000	$4,546	$69	$1,965	$51,580
5	85.55	611	250	33	$30,000	$5,671	$2,836	$45,000	$4,547	$122	$1,966	$51,635
5	95.37	814	250	33	$30,000	$7,565	$3,783	$45,000	$4,547	$163	$1,965	$51,675

TABLE G-1. (Continued)

CASE 3 8785.7 1000

TOTAL INSTALLED DIRECT CAPITAL	ANNUAL COST FOR CAPITAL RECOVERY	Blower HP	ACcb (carbon)	ACe (elect.)	Maint. Hours	ACm (Maint.)	ACol (Op. Labor)	Steam Required (lb/regen)	Solvent Recovered	ACp (credit for solvent)	ACs (steam)	ACmem (membrane replacement)	ACcw (cooling water)
$5,058,426	$823,006	2.01	$790	$468,913	79.4	$3,784	$9,785	557	75.70	($38,836)	$10.052	$1,031,359	$4,898
$9,231,602	$1,501,982	3.27	$1,156	$469,354	81.1	$3,864	$9,785	814	107.44	($55,115)	$14,697	$1,953,828	$7,162
$13,157,714	$2,140,760	4.11	$1,999	$469,648	82.6	$3,939	$9,785	1408	120.13	($61,628)	$25,421	$2,814,655	$12,387
$958,480	$155,945	2.18	$815	$468,974	79.7	$3,799	$9,785	574	75.73	($38,848)	$10,360	$119,470	$5,048
$1,440,897	$234,434	3.50	$1,189	$469,436	81.5	$3,887	$9,785	837	107.18	($54,982)	$15,120	$221,326	$7,368
$1,927,663	$313,631	7.56	$4,068	$470,856	83.2	$3,969	$9,785	2865	120.14	($61,631)	$51,729	$315,325	$25,206
$622,977	$101,358	2.55	$840	$469,103	80.6	$3,842	$9,785	592	75.73	($38,849)	$10,683	$42,504	$5,206
$837,230	$136,217	4.03	$1,850	$469,620	83.0	$3,957	$9,785	1303	107.71	($55,254)	$23,527	$75,617	$11,464
$982,300	$159,820	4.79	$2,063	$469,889	84.8	$4,045	$9,785	1452	120.07	($61,598)	$26,229	$100,872	$12,781

CASE 4 219.6 100

TOTAL INSTALLED DIRECT CAPITAL	ANNUAL COST FOR CAPITAL RECOVERY	Blower HP	ACcb (carbon)	ACe (elect.)	Maint. Hours	ACm (Maint.)	ACol (Op. Labor)	Steam Required (lb/regen)	Solvent Recovered	ACp (credit for solvent)	ACs (steam)	ACmem (membrane replacement)	ACcw (cooling water)
$197,582	$32,147	0.03	$13	$11,714	77.5	$3,697	$9,785	9	0.39	($198)	$45	$25,784	$22
$301,400	$49,038	0.04	$25	$11,720	77.6	$3,699	$9,785	17	0.55	($280)	$65	$48,846	$32
$398,276	$64,799	0.05	$35	$11,724	77.6	$3,701	$9,785	25	0.61	($313)	$77	$70,366	$37
$95,001	$15,457	0.03	$15	$11,715	77.6	$3,697	$9,785	11	0.39	($198)	$45	$2,987	$22
$106,503	$17,328	0.05	$28	$11,721	77.6	$3,700	$9,785	20	0.55	($280)	$67	$5,533	$32
$117,114	$19,055	0.06	$39	$11,726	77.6	$3,702	$9,785	28	0.61	($314)	$79	$7,883	$38
$86,362	$14,051	0.03	$21	$11,717	77.6	$3,699	$9,785	15	0.39	($198)	$48	$1,063	$24
$90,142	$14,666	0.06	$38	$11,725	77.6	$3,701	$9,785	27	0.55	($281)	$73	$1,890	$35
$93,023	$15,135	0.07	$50	$11,729	77.7	$3,704	$9,785	35	0.61	($313)	$84	$2,522	$41

TABLE G-1. (Continued)

CASE 3 8785.7 1000

ACww (waste water)	TOTAL ANNUAL OPERATING COST	ANNUAL COST FOR CAPITAL RECOVERY	TOTAL ANNUAL COSTS	COST PER METRIC TON CONTROLLED
$14,694	$1,505,438	$823,006	$2,328,444	$8,244
$21,485	$2,426,216	$1,501,982	$3,928,197	$9,800
$37,162	$3,313,368	$2,140,760	$5,454,128	$12,169
$15,144	$594,544	$155,945	$750,489	$2,656
$22,103	$695,230	$234,434	$929,664	$2,325
$75,619	$894,925	$313,631	$1,208,555	$2,696
$15,617	$518,731	$101,358	$620,090	$2,195
$34,393	$574,960	$136,217	$711,177	$1,770
$38,342	$602,407	$159,820	$762,227	$1,701

CASE 4 219.6 100

ACww (waste water)	TOTAL ANNUAL OPERATING COST	ANNUAL COST FOR CAPITAL RECOVERY	TOTAL ANNUAL COSTS	COST PER METRIC TON CONTROLLED
$65	$50,927	$32,147	$83,074	$57,821
$95	$73,985	$49,038	$123,023	$60,335
$112	$95,524	$64,799	$160,324	$70,319
$65	$28,133	$15,457	$43,589	$30,329
$97	$30,683	$17,328	$48,011	$23,603
$115	$33,054	$19,055	$52,108	$22,854
$71	$26,229	$14,051	$40,280	$28,026
$106	$27,072	$14,666	$41,738	$20,419
$122	$27,723	$15,135	$42,857	$18,807

TABLE G-1. (Continued)

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alpha*	% Removal	Membrane Area (ft^2)	ACFM to Compr.	Req'd HP for Compr.	Installed Capital Cost	Membrane Capital Cost	Auxil. Equip. Cost	DCCa (ads. vessels)	DCCd (duct work)	DCCc (carbon)	DCCf (fans)	TOTAL ADSORBER DIRECT CAPITAL
200	60.13	83269	2500	334	$128,000	$773,519	$386,760	$124,133	$4,546	$430	$1,965	$131,075
200	85.33	157747	2500	334	$128,000	$1,465,371	$732,685	$127,972	$4,548	$803	$1,966	$135,290
200	95.41	227247	2500	334	$128,000	$2,110,991	$1,055,495	$131,491	$4,548	$1,146	$1,965	$139,149
20	60.15	9646	2500	334	$128,000	$89,602	$44,801	$124,795	$4,546	$494	$1,965	$131,801
20	85.12	17869	2500	334	$128,000	$165,994	$82,997	$129,020	$4,548	$905	$1,966	$136,439
20	95.42	25458	2500	334	$128,000	$236,494	$118,247	$132,856	$4,547	$1,279	$1,965	$140,648
5	60.15	3432	2500	334	$128,000	$31,878	$15,939	$126,843	$4,546	$693	$1,965	$134,047
5	85.55	6105	2500	334	$128,000	$56,713	$28,356	$132,298	$4,547	$1,225	$1,966	$140,036
5	95.37	8144	2500	334	$128,000	$75,654	$37,827	$136,406	$4,547	$1,627	$1,965	$144,545

CASE 6 8785.7 100

alpha*	% Removal	Membrane Area (ft^2)	ACFM to Compr.	Req'd HP for Compr.	Installed Capital Cost	Membrane Capital Cost	Auxil. Equip. Cost	DCCa (ads. vessels)	DCCd (duct work)	DCCc (carbon)	DCCf (fans)	TOTAL ADSORBER DIRECT CAPITAL
200	60.13	333076	10000	1337	$263,000	$3,094,077	$1,547,038	$137,361	$4,546	$1,721	$1,965	$145,593
200	85.33	630986	10000	1337	$263,000	$5,861,483	$2,930,741	$152,480	$4,548	$3,213	$1,966	$162,207
200	95.41	908989	10000	1337	$263,000	$8,443,964	$4,221,982	$188,792	$4,548	$6,877	$1,965	$202,181
20	60.15	38583	10000	1337	$263,000	$358,409	$179,204	$139,978	$4,546	$1,977	$1,965	$148,466
20	85.12	71477	10000	1337	$263,000	$663,977	$331,988	$156,582	$4,548	$3,621	$1,966	$166,717
20	95.42	101834	10000	1337	$263,000	$945,974	$472,987	$196,569	$4,547	$7,677	$1,965	$210,758
5	60.15	13727	10000	1337	$263,000	$127,512	$63,756	$148,046	$4,546	$2,773	$1,965	$157,331
5	85.55	24420	10000	1337	$263,000	$226,851	$113,425	$193,397	$4,547	$7,350	$1,966	$207,260
5	95.37	32577	10000	1337	$263,000	$302,617	$151,309	$216,566	$4,547	$9,761	$1,965	$232,839

TABLE G-1. (Continued)

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TOTAL INSTALLED DIRECT CAPITAL	ANNUAL COST FOR CAPITAL RECOVERY	Blower HP	ACcb (carbon)	ACe (elect.)	Maint. Hours	ACm (Maint.)	ACol (Op. Labor)	Steam Required (lb/regen)	Solvent Recovered	ACp (credit for solvent)	ACs (steam)	ACmem (membrane replacement)	ACcw (cooling water)
$1,419,353	$230,929	0.26	$133	$117,142	78.0	$3,717	$9,785	93	3.85	($1,976)	$446	$257,840	$217
$2,461,346	$400,461	0.42	$248	$117,200	78.4	$3,737	$9,785	174	5.47	($2,804)	$649	$488,457	$316
$3,433,636	$558,653	0.55	$353	$117,244	78.8	$3,756	$9,785	249	6.11	($3,135)	$766	$703,664	$373
$394,204	$64,137	0.27	$152	$117,148	78.0	$3,721	$9,785	107	3.85	($1,976)	$446	$29,867	$217
$513,431	$83,535	0.45	$279	$117,211	78.5	$3,743	$9,785	196	5.45	($2,797)	$665	$55,331	$324
$623,389	$101,425	0.59	$394	$117,258	78.9	$3,763	$9,785	278	6.11	($3,135)	$788	$78,831	$384
$309,864	$50,415	0.34	$214	$117,171	78.3	$3,732	$9,785	150	3.85	($1,976)	$483	$10,626	$235
$353,105	$57,450	0.55	$378	$117,247	78.9	$3,760	$9,785	266	5.48	($2,811)	$727	$18,904	$354
$386,026	$62,806	0.69	$501	$117,294	79.3	$3,782	$9,785	353	6.11	($3,133)	$836	$25,218	$407

CASE 6 8785.7 100

TOTAL INSTALLED DIRECT CAPITAL	ANNUAL COST FOR CAPITAL RECOVERY	Blower HP	ACcb (carbon)	ACe (elect.)	Maint. Hours	ACm (Maint.)	ACol (Op. Labor)	Steam Required (lb/regen)	Solvent Recovered	ACp (credit for solvent)	ACs (steam)	ACmem (membrane replacement)	ACcw (cooling water)
$5,049,708	$821,588	0.82	$530	$468,496	79.4	$3,784	$9,785	373	7.57	($3,884)	$1,221	$1,031,359	$595
$9,217,432	$1,499,676	1.38	$990	$468,692	81.1	$3,864	$9,785	697	10.74	($5,511)	$1,796	$1,953,828	$875
$13,131,127	$2,136,434	1.81	$2,119	$468,843	82.6	$3,939	$9,785	1492	12.01	($6,163)	$3,129	$2,814,655	$1,525
$949,079	$154,415	0.88	$609	$468,519	79.7	$3,799	$9,785	429	7.57	($3,885)	$1,221	$119,470	$595
$1,425,683	$231,959	1.50	$1,116	$468,735	81.5	$3,887	$9,785	786	10.72	($5,498)	$1,861	$221,326	$907
$1,892,720	$307,946	1.97	$2,366	$468,899	83.2	$3,969	$9,785	1666	12.01	($6,163)	$3,254	$315,325	$1,586
$611,598	$99,507	1.13	$855	$468,606	80.6	$3,842	$9,785	602	7.57	($3,885)	$1,367	$42,504	$666
$810,536	$131,874	1.87	$2,265	$468,864	83.0	$3,957	$9,785	1595	10.77	($5,525)	$3,039	$75,617	$1,481
$949,765	$154,527	2.36	$3,008	$469,038	84.8	$4,045	$9,785	2118	12.01	($6,160)	$3,535	$100,872	$1,723

TABLE G-1. (Continued)

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ACww (waste water)	TOTAL ANNUAL OPERATING COST	ANNUAL COST FOR CAPITAL RECOVERY	TOTAL ANNUAL COSTS	COST PER METRIC TON CONTROLLED
$652	$387,955	$230,929	$618,884	$43,076
$949	$618,537	$400,461	$1,018,998	$49,975
$1,120	$833,927	$558,653	$1,392,579	$61,080
$652	$160,012	$64,137	$224,149	$15,596
$972	$185,514	$83,535	$269,049	$13,227
$1,152	$209,221	$101,425	$310,647	$13,624
$706	$140,975	$50,415	$191,390	$13,317
$1,063	$149,408	$57,450	$206,858	$10,120
$1,221	$155,911	$62,806	$218,718	$9,598

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ACww (waste water)	TOTAL ANNUAL OPERATING COST	ANNUAL COST FOR CAPITAL RECOVERY	TOTAL ANNUAL COSTS	COST PER METRIC TON CONTROLLED
$1,784	$1,513,670	$821,588	$2,335,257	$82,681
$2,626	$2,436,945	$1,499,676	$3,936,621	$98,210
$4,574	$3,302,405	$2,136,434	$5,438,839	$121,348
$1,785	$601,897	$154,415	$756,312	$26,769
$2,720	$704,838	$231,959	$936,797	$23,428
$4,757	$803,776	$307,946	$1,111,722	$24,803
$1,999	$525,740	$99,507	$625,247	$22,130
$4,442	$563,925	$131,874	$695,799	$17,315
$5,168	$591,014	$154,527	$745,541	$16,642